Experimental and Computational Studies on Fluidized Bed Biomass Gasifier for Production of Clean Energy

Dissertation submitted to the

National Institute of Technology Rourkela

in partial fulfillment of the requirements

of the degree of

Doctor of Philosophy

In

Chemical Engineering

By

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January, 2016

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January, 2016

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This is to certify that the thesis entitled "***Experimental and Computational Studies on Fluidized Bed Biomass Gasifier for Production of Clean Energy***"submitted by **Mr. *Deo Karan Ram*** (Roll No. **511CH105**) to National Institute of Technology, Rourkela towards partial fulfillment of the requirements for the award of the***Doctorof Philosophy*** degree in Chemical Engineering, is a bonafide record of his work carried out under my supervision and guidance.

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Declaration of Originality

I, *DEO KARAN RAM*, Roll Number *511CH105* hereby declare that this dissertation entitled "“***Experimental and Computational Studies on Fluidized Bed Biomass Gasifier for Production of Clean Energy***”"represents my original work carried out as a doctoral student of NIT Rourkela and, to the best of my knowledge, it contains no material previously published or written by another person, nor any material presented for the award of any other degree or diploma of NIT Rourkela or any other institution. Any contribution made to this research by others, with whom I have worked at NIT Rourkela or elsewhere, is explicitly acknowledged in the dissertation. Works of other authors cited in this dissertation have been duly acknowledged under the section "Bibliography". I have also submitted my original research records to the scrutiny committee for evaluation of my dissertation.

I am fully aware that in case of any non-compliance detected in future, the Senate of NIT Rourkela may withdraw the degree awarded to me on the basis of the present dissertation.

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ACKNOWLEDGEMENTS……..

In pursuit of this academic endeavor, I feel that I have been singularly fortunate; as inspiration, guidance, direction, cooperation, love and care all came in my way in abundance and it seems almost an impossible task for me to acknowledge the same in adequate terms. Yes, I shall be failing in my duty if I don't record my profound sense of indebtedness and heartfelt gratitude to my supervisor Dr. (Mrs.) AbantiSahoo who guided and inspired me in pursuance of this work. She has not only guided in just technical matters but has always taught several important points to gain maturity to work and her nature of thinking has always influenced me in many ways. Her association will remain a beacon light to me throughout my career.

I owe a depth of gratitude to Prof P. Rath, Head, Department of Chemical Engineering for providing the necessary facilities for the project work. I would like to express my gratitude to my Doctoral Scrutiny Committee (DSC) Dr. H. M Jena, Dr. A. Kumar, Civil Engineering Department and Dr. K.K. Khatua of Civil Engineering Department for their thoughtful advices given during discussion sessions. I would like to thank the all the faculties of Chemical Engineering Department for their support throughout my research work.

I am also thankful to **Ministry of New Renewable Energy, Govt. of India, New Delhi** for sanctioning the project to my supervisor by which a Fluidized bed Biomass Gasifier was installed in the Chemical Engineering Department of NIT Rourkela and I got the chance to carry out my PhD work.

I want to acknowledge the support from all non-teaching staff members and friends in Chemical Engineering Department. I would like to greatly acknowledge and thank the entire Administration and Management of National Institute of Technology, Rourkela, for enabling and supporting me for this work.

Finally, I would like to owe a deep sense of thankfulness to all my family members particularly to my parents for their much appreciated support, encouragement and best wishes for my studies. Last but definitely not least, I am really grateful to almighty for those joyful moments I enjoyed and painful instances which made me tough and strong to face situations in life to come and for the exceptional journey and memories at National Institute of Technology Rourkela.

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Contents

List of Figures

List of Tables

Nomenclature

a	Speed of Sound
C_D	Drag Co-efficient
$C_{fr,ls}$	Coefficient of Friction Between the l^{th} and s^{th} Solid Phase Particles
$C_{i\varepsilon}$	Constants
d	Diameter
e_{ls}	Coefficient of Restitution
F	Force
$\vec{g}$	Acceleration due to Gravity
g_0	Radial Distribution Function
$g_{0,ls}$	Radial Distribution Co-efficient
G_b	Generation of Turbulence Kinetic Energy due to Buoyancy
G_k	Generation of Turbulence Kinetic Energy due to the Mean Velocity Gradients
$G_{k,q}$	Turbulence Kinetic Energy
h	Specific Enthalpy
I_{2D}	Second Invariant of the Deviatoric Stress Tensor
K	Rate Constant
K_{ls}	The Fluid-solid and Solid-solid Exchange Coefficient
K_{pq}	Interphase Momentum Co-efficient
K_{Θ_s}	Diffusion Co-efficient
M	Mash Number
N	Total Number of Phases
p	Pressure
Pr_t	Turbulent Prandtl Number
q	Heat Flux
R	Rate of Reaction
S_k, S_ε	User-defined Source Terms
S_q	Source Term
T	Temperature
$\vec{U}_q$	Phase-weighted Velocity
V	Volume
v	Velocity
v'	Stoichiometric coefficient of reactant
v''	stoichiometric coefficient of product
Y_i	Mass Fraction of Species
Y_M	Contribution of the Fluctuating Dilatation in Compressible Turbulence to the Overall Dissipation Rate
Υ_{Θ_s}	Collisional Dissipation of Energy

Greek letters

α	Volume Fraction
β	Coefficient of Thermal Expansion
∇	Gradient
ε_q	Dissipation Rate
η	Rate Exponent
Θ_s	Solid Phase Granular Temperature
λ_s	Bulk Viscosity
μ	Viscosity
μ_s	Solid Shear Viscosity
$\mu_{s,col}$	Collision Viscosity
$\mu_{s,fr}$	Frictional Viscosity
$\mu_{s,kin}$	Kinetic Viscosity
$\Pi_{k_q} = \Pi_{\varepsilon_q}$	Influence of Dispersed Phase on Continuous Phase q
ρ	Density of Fluid
ρ_m	Density of Bed materials
σ_k	Turbulent Prandtl Numbers For k
σ_ε	Turbulent Prandtl Numbers For ε
Γ_ϕ	Diffusion Co-efficient Forϕ
$\bar{\bar{\tau}}$	Stress-strain Tensor
$\tau_{F,pq}$	Characteristic Relaxation Time
τ_p	Particulate Relaxation Time
ϕ	Angle of Internal Friction
Φ_{ls}	Energy Exchange Between l^{th} Solid Phase and s^{th}solid Phase

Subscripts

j	Species
p, q	Phase
r	Reaction
s	Solids

Abbreviations

2-D	Two Dimensional
3-D	Three Dimensional
CFD	Computational Fluid Dynamics
E/R	Equivalence ratio
FB	Fluidized Bed
FVM	Finite Volume Method
GAMBIT	Geometry and Mesh Building Intelligent Toolkit

KTGF	Kinetic Theory Of Granular Fluid Bed
PDE	Partial Differential Equations
Re	Reynolds Number
S/B	Steam to biomass ratio
SIMPLE	Semi-implicit Method for Pressure-linked Equations
TFM	Two Fluid Models
CH_4	Methane
CO	Carbon Monoxide
CO_2	Carbon Dioxide
H_2	Hydrogen
H_2O	Water
N_2	Nitrogen
O_2	Oxygen

Abstract

An energy efficient approach to hydrogen rich syn-gas production from biomass using a fluidized bed Gasifier is presented. A fluidized bed gasifier is designed and installed in the laboratory by fabricating outside in parts. The effects of different biomass materials, temperature, steam to biomass ratio (S/B) and Equivalence Ratio (ER) on gas yield, gas composition, and carbon conversion efficiency have been studied. Catalytic effects are also studied by changing the bed materials (viz. sand, dolomite, red mud and their mixtures). Different biomass samples such as rice husk, rice straw, saw dust, wood chips, sugarcane bagasse and coconut coir have been gasified in the present work with different bed materials. Temperature during gasification was varied with 500-1000^{0}C. ER was varied within 0.15 to 0.35 and steam to biomass ratio was varied within 1.35 to 2.5. Attempt is made to develop correlation for the yield of hydrogen on the basis of dimensional analysis by relating different system parameters for all the biomass feed samples. Carbon conversion efficiency was observed to vary within 70 – 97%. Experimental results show hydrogen yields to vary within 56-74 gm per kg of feed sample for different biomass samples. The calculated values of H_2yieldare compared against the experimentally observed data. It is observed that higher temperature contributes to higher gas yield and higher carbon conversion. CFD simulations have also been carried out for optimization of process parameters. The gas-solid interaction, the thermal-flow behavior and gasification process inside a fluidized-bed biomass gasifier are studied using the commercial CFD solver ANSYS/FLUENT15.0. A 2-D and 3-D model based on Eulerian-Eulerian approach coupled with granular kinetic theory has been developed to simulate the bed hydrodynamics and heat transfer for the FBG where volume fraction, bed pressure drop, temperature profile have been focused using FLUENT software. The influences of particle properties viz. gas velocity and temperature of bed material within the gasifier have been investigated comprehensively for simulation which provides a powerful basis for accurate design of FBG. Simulation and experimental observations are found to have very good approximation in most of the cases thereby validating the results against each other. Performance of fluidized bed gasifier is found to be satisfactory for ER = 0.25, S/B = 0. 5 and temperature=700^{o}C with the use of red mud sand mixture as bed material. Little deviation among carbon conversion efficiency and thermal efficiency for all samples ensure that this technology can be used successfully for clean energy production and the developed correlations can be used for other biomass samples over a wide range of parameters.

Keywords:-Gasification, Syn-gas analysis, Energy balance, Carbon conversion efficiency, Cold gas efficiency and energy analysis, CFD.

Chapter-1

Introduction

With continuously increasing demand for energy, our current primary energy source, fossil fuels are getting depleted to support the economic growth. As a result there is a significant impact on the global climate change. There is also concern for the availability of the fossil fuels in the near future for which the price of fossil fuels is fluctuated. Now a reliable and sustainable energy supply has been a major concern for the global community.

1.1 Background

To respond the energy crisis it has become essential not only to use the existing energy sources efficiently but also to develop alternative or non-conventional sources of energy. In this context a lot of effort has been made to explore renewable energy production technologies around the world such as hydroelectric, geothermal, wind, solar and biomass. Of the various renewable energy sources available, biomass appears to offer a promising solution to tackle the ever increasing energy demand [1]. Biomass energy products are generated from agricultural crops and residues, herbaceous and woody materials and organic wastes. These materials can either be directly combusted for energy production or processed into energy products which are then used as transportation fuels or for the production of electricity and heat. Biomass is formed through photosynthesis where sunlight is converted to chemical energy. Such chemical energy is stored in chemical bonds of plant. During gasification, when the biomass feed samples are subjected to high temperature, the chemical bonds among Carbon, Hydrogen and Oxygen molecules present in plants are broken and stored energy is released [2].

In developing countries like India, biomass in its natural form in dry condition is still widely used in rural communities as a major heat source, setting aside a small part of it form manure and compost. In this respect too, biomass is used very selectively and a large variety of biomass is allowed to perish in the environment. Leaving aside the selective type of biomass which is lumpy and smokeless and used for domestic heating purposes, light grainy and powdery biomass finds no use. Moreover, their uses in industrial sector is very limited as they occur in a scattered manner and are not collected in an organized way due to their bulk volume and low end-value. Although biomass is not a major industrial fuel, it supplies 15–20% of the total fuel use in the world. It is used mostly in non-industrialized economies for domestic heating and cooking. In industrialized countries, the use of biomass as a fuel is

largely restricted to the use of by-products from forestry and the paper & sugar industries [3]. India produces about 420 million tons of biomass (fire wood: 220 million tons and agro wastes including powdery biomass around 220 million tons). Considering the vast availability, it may be expected that small industrial installation with an appropriate technology may either add value to it or exploit efficiently the potential heat energy present in it. Relatively new technologies have come up to help in this respect. Biomass, instead of being used in solid form directly is converted for use in gaseous form through gasification route. Production of hydrogen from renewable biomass has several advantages compared to that of fossil fuels. A number of processes are being practiced for efficient and economic conversion and utilization of biomass to hydrogen [4].

A wide variety of biomass can be converted to energy by using gasification. Biomass can either be produced from wastes which are discarded having no apparent value or dedicated energy crops can specifically be grown for the production of bioenergy. Gasification is a process that converts organic or fossil based carbonaceous material into gaseous fuel through partial oxidation. Of the various renewable energy sources available, biomass appears to offer a promising solution to tackle the ever increasing energy demand and biomass energy ensures the sustainability of energy supply in the long term by reducing the impact on the environment. Biomass has a lower carbon footprint and do not contribute to overall carbon emissions. In addition the biomass gasification offers the advantage of using wastes and residues, improved land management for agriculture and forest [5]. Thermochemical gasification of biomass is a well-known technology that seems to be a feasible application that has been developed for industrial applications [6-9]. The researchers are recommending an alternative fuel and efficient conversion techniques to overcome the problems of energy crisis and environmental damages. Biomass gasifier is a device used to generate gas at a lower price than other fuels [10]. When biomass energy is obtained by using agricultural waste as fuel, it is considered as "CO_2 neutral" because emissions of sulfur dioxides and nitrogen oxides are very small. Thus, the use of agricultural waste is a real option for clean fuel with zero emissions.

It is always convenient and economical to burn the solid, semi-dried biomass and obtain useful heat at the location of biomass source. The heat derived from the combustion of biomass can be used for several useful processes such as cooking, industrial heat requirements, steam generation, production of electrical energy from steam, etc. However, when the energy is to be transported over a long distance, it is more economical to convert the biomass into liquid or gaseous fuels and then transport them through pipeline or by tanks and then use the fuels in liquid or gaseous forms at the receiving end. Alternatively the biomass is converted to electrical energy in a biomass thermal, electrical power plant and the

energy is transmitted to electric power to the load center [11].The applications of biomass combustion process cover a broad range of ratings from a fraction of kilowatt (for cooking) to a few megawatts (in municipal waste-to-energy electrical power plant).

Hydrogen is one of the potential carriers of energy that could be used to replace the existing fossil fuels. Besides the zero carbon footprints, hydrogen is expected to become a prominent energy carrier for stationary and mobile power generation applications such as in transport, industrial, commercial, and residential applications [12]. The utilization of renewable sources including the biomass of forestry, agricultural, and municipal waste has become a new source of energy due to the abundance of these wastes. Consequently, producing hydrogen from biomass not only offers a zero net carbon emission but also generates electricity and heat which is clean. Biomass gasification is considered as one of the potential alternatives for the production of hydrogen, a clean energy.

1.2 Advantages of Biomass Gasification

Usually, conventional gasification refers to coal gasification. But use of biomass sample for gasification is now getting importance because of many advantages like less time requirement for biomass conversion, low emission of gaseous pollutants (SOx & NOx), and production of small density char in comparison with that of conventional coal gasification.

In the gasification process the organic matters are converted into fuels known as syngas at high temperature and in a controlled environment in the presence of steam. Syngas is a type of an effective fuel. The process of gasification has helped the industry to utilize organic material to generate electricity and helps the industrial plants to reduce their production cost. The recent development in the gasification process has drawn the attention of industry to use plastic as a combustion material. The syngas generated in the process of gasification is used to produce electricity and effective mechanical power. As compared to the solid fuels, gaseous fuel is believed to be more environments friendly. The process of gasification does not emit greenhouse gases in the air.

The electric power generated in this process is much cheaper than the steam cycle. The increasing use of this process has also attracted the automobile industry to make cars that can use syngas as a fuel. Now a day the use of gasification is also popular in agriculture. Gasification is a vital process to save the major fertilizer and chemical industry [1].

1.3 Types of Gasifiers

Depending upon the gasification medium, gasifiers can be broadly classified into two groups:

1. Air-blown, where air is the gasification medium

2. Oxygen-blown, where pure oxygen is the gasification medium

As there is an interaction of air/oxygen and biomass in the gasifier, they are classified according to the way air/oxygen is introduced into it. Depending upon how the gas and fuel contact each other, gasifiers are further divided into following four types:

1. Entrained bed
2. Fluidized bed (Bubbling or Circulating)
3. Spouted bed
4. Fixed or moving bed

Depending upon the flow pattern fluidized bed gasifiers are divided into two major types.

1. Bubbling fluidized bed gasifier
2. Circulating fluidized bed gasifier

1.4 Fluidized Bed Gasifier

Fluidized bed systems are developed to provide uniform temperatures and efficient contact between gases and solids and to minimize the formation of hot spots within the gasifier. A fluidized bed gasifier is suitable for small to medium sized particles which are fluidized by a suitable gasification medium such as air or steam. The fluidized bed contains either inert material, sand or reactive materials, limestones or catalysts and the materials are kept in suspension by a rising column of gas which can be air, oxygen or steam. The fuel is fed into either a suspended or a circulating fluidized bed. This is where the fuel particles mix very quickly, providing a high heat transfers rate and rapid pyrolysis. Compared to fixed–bed gasifiers, the temperatures are lower in fluidized bed gasifiers (750^{o} C-900^{o} C). In fixed bed gasifiers temperature ranges within1000^{o}C-1200^{o}C. The fluidized bed gasifiers are being widely used by many researchers for hydrogen production from biomass samples [13].

1.4.1 Advantages of Fluidized Bed Gasification

The fluidized bed gasification process has several advantages compared to simple burning process and other forms of gasification. Some of these advantages are described below:

1. It is highly efficient as the overall thermal efficiency of fluidized bed gasifiers is typically in the range of 75% to over 90%, depending on the ash and moisture content of the fuel.
2. In this gasifier air to fuel ratio can be changed which also helps to control the bed temperature in addition to the yield.
3. Fluidized bed gasifiers are more tolerant to variation in feedstock as compared to other types of gasifiers.
4. Such gasifiers maintain uniform radial temperature profiles and avoid slugging problems.

5. Higher throughput of fuel as compared to other gasifiers.

Fluidized bed gasifier has capacity of Flexible Operations, because the process produces a fuel gas rather than just quantities of heat, which can be easily applied to a variety of industrial processes including boilers, dry kilns, veneer dryers or several pieces of equipment at once.

1.4.2 Disadvantages of Fluidized Bed Gasification

1. Oxidizing conditions are created when oxygen diffuses from bubble to the emulsion phase there by reducing the gasification efficiency.
2. Reduced solid conversion due to intimate mixing of fully and partially gasified fuels.
3. Losses occurring due to particle entrainment.

1.4.3 Bubbling fluidized bed gasifier

Bubbling fluidized-bed gasifiers contain fine inert particles of sand or alumina. These inert particles break up the biomass samples fed into the bed to ensure proper heat transfer. As gas is forced through the inert particles, a point is reached when the frictional forces between the particles and the gas counter balance the weight of the fluid. A disadvantage of bubbling fluidized –bed gasification is that the formation of large bubbles may result in some gas-bypassing through the bed. The advantages of bubbling fluidized-bed gasification include the following:

- Yields a uniform product gas
- Exhibits uniform temperature distribution throughout the gasifier
- Able to accept a wide range of biomass particle sizes, including fines
- Provides high rates of heat transfer between inert materials , fuel, and gas
- High conversion possible with low amounts of tar and unconverted carbon.

1.4.4 Circulating fluidized bed gasifier

In a circulating fluidized –bed gasifiers, high gas velocities result in an entrainment of some particles, which escape from the top of the gasifier vessel. The entrained particles are separated in a cyclone and returned to the reactor. The advantages of circulating fluidized-bed gasification are as follows:

- Suitable for rapid reactions
- High heat transfer rates possible due to high heat capacity of bed material
- High conversion rates possible with low amounts of tar and unconverted carbon.

The disadvantages of circulating fluidized-bed gasification include the following:

- Temperature gradients occur in direction of solid flow
- Size of fuel particles determines minimum transport velocity and high velocities may result in equipment erosion
- Less efficient heat exchange than bubbling fluidized bed.

The concern for climate change has increased the interest in biomass gasification for which fluidized bed gasifiers are particularly popular, occupying nearly 20% of their market. Due to the above mentioned disadvantages of a circulating fluidized bed, a bubbling fluidized bed gasifier is selected for the present study.

1.5 Computational Fluid Dynamics

Computational fluid dynamics (CFD) is one of the branches of fluid mechanics that uses numerical methods and algorithms to solve and analyze problems that involve fluid flows. Fluid flows are governed by partial differential equations (PDE) which represent conservation laws for the mass, momentum and energy. Computational Fluid Dynamics (CFD) is used to replace such PDE systems by a set of algebraic equations which can be solved using digital computers. Due to a combination of increased computer efficacy and advanced numerical techniques, the numerical simulation techniques such as CFD becomes a reality and offers an effective means of quantifying the physical and chemical process in the biomass thermo- chemical reactors under various operating conditions within a virtual environment. The results of accurate simulations can help to optimize the system design and operation and understand the dynamic process inside the reactors. CFD modeling techniques are becoming widespread in the biomass thermo-chemical conversion areas. Researchers have been using CFD to simulate and analyze the performance of thermo-chemical conversion equipment such as fluidized beds, fixed beds, combustion furnaces, firing boilers, rotating cones and rotary kilns. CFD programs predict not only fluid flow behavior, but also heat and mass transfer, chemical reactions (e.g. devolatilization, combustion), phase changes (e.g. vapour in drying, melting in slagging), and mechanical movement (e.g. rotating cone reactor). Compared to the experimental data, CFD modeling results are capable of predicting qualitative information and in many cases accurate quantitative information. CFD modeling has established itself as a powerful tool for the development of new ideas and technologies [14]. CFD provides a qualitative prediction of fluid flows by means of

1. Mathematical modeling (partial differential equations)
2. Numerical methods (discretization and solution techniques)
3. Software tools (solvers, pre- and post-processing utilities)

1.6 Overview of the Project Topic

Hydrogen obtained through biomass gasification is therefore currently considered as a clean and most promising source of energy. It is very difficult and also very much time consuming to get the exact optimum conditions for a fluidized bed gasifier through experimentations. Trying several times on trial and error basis by varying different parameters is also very much power consuming. Sometimes carrying out experiments might not be economically viable at all. Therefore CFD modelling has proven to be a viable option over recent years. With the continual enhancement of computational capabilities, it is possible to carry out any modification to the optimum design or in operating conditions before actual experimentations. Very little literature is found on CFD modelling for FBG. Therefore, it is thought to carry out CFD modelling for the hydrodynamic studies, thermal flow behavior existing inside the fluidized bed gasifier along with some experimental investigations.

Chapter-II

Literature Review

It has now become essential not only to use the existing energy sources efficiently but also to develop alternative or non-conventional sources of energy. Gasification is a process that converts solid biomass efficiently into combustible gas (i.e. mixture of CO, CH_4 and H_2), with char, water and condensable as minor products. Biomass energy ensures the sustainability of energy supply in the long term by that reduces the impact on the environment. Fluidized Bed Gasifier (FBG) can handle all types of dry, small sized biomass wastes. It can be operated in both batch and continuous mode. FBG handling biomass produces syngas with high calorific value and solid wastes with less ash content. Thus, wastes from agro- industry can also be used for power generation with proper gasification technology.

2.1 Energy Route

It is always convenient and economical to burn the solid, semi-dried biomass and obtain useful heat at the location of biomass source (e.g. sugar cane bagasse can be burnt near a sugar factory site). The energy route of the combustion process is as explained below [5, 15, 16, 17].

$$\text{(Dry Shredded Biomass)} \rightarrow \text{Burning} \rightarrow \text{Heat of Combustion} \quad \uparrow \quad \text{Air} \tag{2.1}$$

Energy route with biomass resources is shown below in block diagrams.

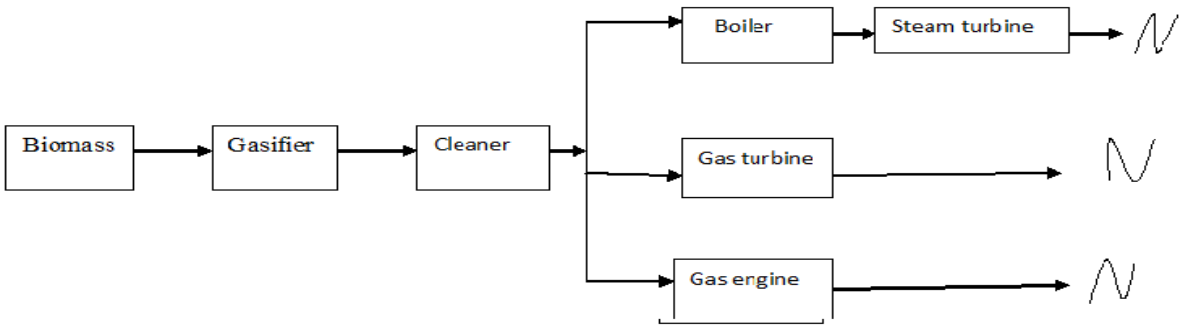


Fig.-2.1(a) : Block diagram for energy production process via biomass gasification route

2.2 Gasification Fundamentals

Gasification is a process that converts organic or fossil-based carbonaceous materials into carbon monoxide, hydrogen, carbon dioxide, methane and nitrogen (if air is used as the oxidizing agent). This is achieved by reactions of the materials at high temperatures with a controlled amount of air, oxygen or steam. It contains a series of steps: drying, devolatilisation (Pyrolysis), char gasification (gas-solid reactions) and gas phase reactions. Also, the final product gas composition is a result of important endothermic and exothermic chemical reactions that take place inside the gasifier. The exothermic reactions provide heat to support the endothermic reactions through partial combustion. Eventually a steady state is reached and the gasifier maintains its operation at a certain temperature. In most of the applications, the gas producer, which is called gasifier, is a simple device consisting of a cylindrical container. The resulting gas mixture is called syngas, synthesis gas or producer gas and is itself a fuel. Syngas can be burned directly in gas engines, internal combustion engines (both compression and ignition), used as a substitute for furnace oil in direct heat applications and can be used to produce methanol in an economically viable way which is used as chemical feedstock for industries. It can also be converted into synthetic fuel via Fischer-Tropsch process.

2.3 Gasifying Mediums

The gasification process requires gasification agent for the thermo-chemical conversion of carbonaceous feed stock. Oxygen, air, steam or a combination of these is used as the oxidizing agent for the requirement of quality of the product gas. When the gasifying agent is air, the process is named air gasification and the producer gas has lower quality in terms of heating value due to the high percentage of nitrogen mixed in the gas. This gas is suitable for boilers, engines and turbines.

If the gasifying agent is pure oxygen or steam, it is called oxygen or steam gasification respectively. In this case the producer gas has relatively higher quality and can be used for conversion to methanol and gasoline. In the present study air is taken as gasifying medium.

2.4 Zones of Gasifier

Gasification process is carried out in different stages or zones as described below and shown in Fig.2.1(b)[1].

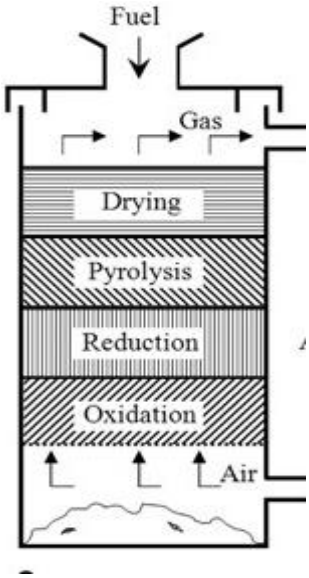


Fig.-2.1 (b) Fluidized Bed Gasifier

2.4.1 Drying Zone

The main operation in drying zone is the removal of moisture. Biomass fuels consist of moisture ranging from 5 to 35%. At the temperature above 100°C, the water is removed and converted into steam. Biomass sample does not experience any kind of decomposition in this zone.

2.4.2 Pyrolysis Zone

Pyrolysis is the thermal decomposition of biomass in the absence of oxygen. The main reaction in this zone is the irreversible devolatilization reaction. Energy required for the reaction is obtained from the oxidation zone and temperature lies in between 200 and 500°C.

Pyrolysis of biomass samples generally produces three types of products:

- Gases like H_2, CO, CH_4, H_2O, and CO_2
- Tar, a black, viscous and corrosive liquid
- Char, a solid residue containing carbon

2.4.3 Oxidation Zone

This zone provides the energy for the gasification process i.e. for drying, pyrolysis and reduction. All these reactions are exothermic in nature [18 & 19]. The combustion takes place within the at temperature range of 800 to 1200°C. Heterogeneous reaction takes place between oxygen in the air and solid carbonized fuel producing carbon dioxide as per the following reaction.

$$C + O_2 = CO_2 \quad (2.2)$$

Hydrogen in fuel reacts with oxygen in the air and blasts producing steam as follows.

$$H_2 + ½\ O_2 = H_2O \quad (2.3)$$

1.4.4 Reduction Zone

In the reduction zone, a number of high temperature chemical reactions take place in the absence of oxygen. The major reactions in this zone are water gas reaction, the water shift reaction, the boudouard reaction and methanation reaction. The fuel in this zone is in the highly carbonized form and red hot with all the volatile matters driven off and the temperature in this zone is in between 600 and 800°C. These reactions are mentioned below.

Water gas reaction

$$C + H_2O = CO + H_2 \quad (2.4)$$

Water shift reaction

$$CO + H_2O = CO_2 + H_2 \quad (2.5)$$

Boudouard reaction

$$C + CO_2 = 2CO \quad (2.6)$$

Methanation reaction

$$C + 2H_2 = CH_4 \quad (2.7)$$

2.5 Mechanism of Fluidized Bed Gasifier

Fluidization is one of the best ways of interacting solid particles with fluids when drag force acts on the solid particle and is equal to gravity force / weight of the particles. The fluidized bed is one of the best known contacting methods used in processing industries. The solid particles are transformed to fluid – like state through the contact with fluid i.e. gas or liquid or both which is allowed to pass through a distributor plate. In the fluidized state, the gravitational force pull on solid particles is offset by the fluid drag force on them, thus the particles remain in a semi – suspended condition. At the critical value of fluid velocity, the upward drag force exerted by solid particles becomes exactly equal to the downward gravitational force, causing the solid particles to be suspended within the fluid. At this critical value, the bed is said to be just fluidized. Thereof the solid particles exhibit behaviours of fluid. This critical velocity is known as minimum fluidization velocity [20]. The different flow regimes resulted in the fluidized bed depending on the fluid flow rate is shown in Fig. 2.2. The major challenge of gasification technology is to improve quality of the product gas

which determines the extent of the post-treatment. Tar formation (complex hydrocarbons, CxHyOz) can put an investment in great risk. Multiphase flow, gas-solid interaction, chemical reactions and turbulence are responsible for the composition of the raw output gas. So far, many empirical models and structures have been developed which fail to optimize the technology and result in industrial-scale units. For this reason, computational fluid dynamic (CFD) simulations are being developed. However, the lack of knowledge in the field of chemical reactions puts a big barrier on the accuracy of the simulation projects.

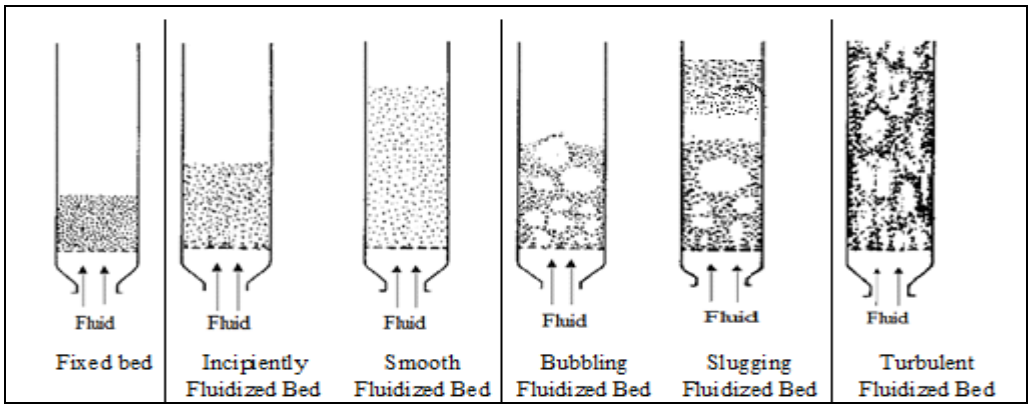


Fig. 2.2 - Flow Regimes of Fluidized Bed

2.6 Gasifier Performance

The gasifier performance can be determined in terms of different efficiencies. These efficiencies are defined [1] as follows.

(a) Gasifier efficiency is defined as the ratio of total energy output to the energy input for carrying out the gasification process for different biomass samples.

(b) Alternately thermal conversion efficiency is calculated as the ratio of the net heating value of the flue gas (NHV) as indicated by Gas analyser to HHV of the biomass sample.

(c) Cold gas efficiency of the gasifier is defined as thetotal energy output with different components of syngas to HHV of the biomass sample.

(d) The carbon conversion efficiency is defined as ratio of carbon content associated with CO, CH_4 and CO_2 in dry product gas to the carbon content present in fuel sample.

2.7 Computational Fluid Dynamics

The basic principle behind CFD modeling method is that the simulated flow region is divided into small cells. Differential equations of mass, momentum and energy balance are discretized and represented in terms of the variables at any predetermined position within/at the center of cell. These equations are solved iteratively until the solution reaches the desired accuracy (ANSYS Fluent 15.0). CFD simulation method is widely used to analyze the fluid

flow behaviours as well as heat and mass transfer processes and chemical reactions. Fluidized bed operations are not economical for small scale applications for which mathematical models are being helpful for designing, for predicting gasifier behavior and studying effect of operating parameters on gasifier performance, startup, shutdown etc.CFD modelling is being used by various researchers in areas specifically in biomass gasification and combustion.

2.7.1 ANSYS FLUENT Software

FLUENT is one of the widely used CFD package. ANSYS FLUENT software contain wide range of physical modeling capabilities which are used to model flow, turbulence, reaction and heat transfer for industrial application. Features of ANSYS FLUENT [21] software are mentioned below.

- **Mesh Flexibility :**
  ANSYS FLUENT software provides mesh flexibility. It has ability to solve flow problems using unstructured mesh. Mesh types which support in FLUENT include quadrilateral, triangular, hexahedral, tetrahedral, polyhedral, pyramid and prism. Due to automatic nature of creating mesh time is saved.
- **Multiphase Flow :**
  It is possible to model different fluids in a single domain in FLUENT.
- **Reaction Flow :**
  Modeling of surface chemistry, combustion as well as finite rate chemistry can be done in FLUENT.
- **Turbulence :**
  It offers a number of turbulence models to study the effect of turbulence in a wide range of flow regimes.
- **Dynamics and Moving Mesh**:
  The users setup the initial mesh and instruct the motion, while FLUENT software automatically changes the mesh to follow the motion instructed.
- **Post-Processing and Data Export :**
  Users can post process their data in FLUENT software creating contours, path lines and vectors among other things to display the data.

The standard k-ε model is employed in this study to simulate the turbulent flow. In the present work, an Eulerian granular multiphase model is adopted where gas and solid phases are all treated as continua, interpenetrating and interacting with each other everywhere in the computational domain.

2.8 Previous Works

Literature survey was carried out for both experimental and computational (CFD)work on biomass gasification. Some the previous works are reported here.

2.8.1 Experimental Investigations

Many researchers started working on hydrogen production from biomass samples in the early eighties. Some of these literatures [11, 13, 22-34] have been reported here. Almost all have analysed yield of hydrogen from biomass with the heating values of producer gas using biomass gasification. Ghani et al. [13] carried out some experiments to analyse the hydrogen production potential from agricultural wastes. Temperature, equivalence ratio, fluidization ratio and static bed heights were varied for coconut shell and palm kernel. Kentaro et al. [23] carried out rice straw gasification with the effect of steam on char reactions. Presence of silica in char showed the catalytic effect on water-gas-shift reaction and char carbonization was observed to be accelerated by increased steam. Thermochemical equilibrium of the reactive system with the additional parameters such as quantity of steam, bed pressure drop and type of biomass was studied by Detournay et al. [24]. Pengmei et al. [28] have modeled a steady state, one dimensional, isothermal two phase, bubbling fluidized bed biomass gasification with the effects of temperature and equivalence ratio. The temperature distribution in the fluidized bed is relatively constant and typically ranges between 700°C and 900°C [15]. The large thermal capacity of inert bed material plus the intense mixing associated with the fluid bed enable this system to handle a much greater quantity and normally, a much lower quality of fuel [35]. The effects of gasifier temperature, steam to biomass ratio and equivalence ratio on gas composition, carbon conversion efficiency and energy conversion efficiency of the product gas were studied by many researchers [36 & 18]. Agglomeration tendencies with some common agricultural residues were analysed in fluidized bed combustion and gasification system [37]. It is observed that the combustion zone temperature is in the order of 900 – 1000^{o}C as in moving bed gasifiers and 800-900^{0}C in fluidized bed gasifiers. The ashes of biomass feed stocks were observed to have ash fusion temperatures in the range of 800^{o}C to 1500^{o}C.

It is observed from literature that only one or two parameters have been studied by several researchers for one set of experiments to investigate the yield of syngas from biomass gasification. The effect of ER (equivalence ratio) and reaction temperature have been investigated on distribution of products and composition of the syngas [38] where increased ER is observed to increase the yield of hydrogen and carbon monoxide. Effect of Nickel catalyst on gasification of Cuban bagasse in a two-stage gasification reactor is observed to increase the yield up to that of equilibrium conditions [39]. The gasification process has been modeled in a simpler way based on the chemical equilibrium considerations [40]. The

gasifier was integrated with the sugarcane mill for parametric study with the verification of literature data and real systems. Proximate and ultimate analyses for the sugarcane bagasse have been carried out for simulation of mass and energy balances for the gasification process [41]. ASPEN PLUS Simulation for Syngas Production from Sugarcane bagasse using a circulating fluidized bed gasifier has also been carried out by some researchers [42] where a rigorous model based on Gibb's minimum free energy method has been used. Osada et al. [43] used activated carbon and Titania supported Ruthenium (RuC and RuTiO2) catalysts in supercritical water for gasification of sugarcane bagasse where complete gasification was achieved at 673^{o}K. Effect of water density was also studied on the yield of the gas products by them. Khan et al. [44, 45] studied Palm Kernel gasification via integrated catalytic adsorption where quicklime was used as the bed materials and Ni was used as the catalyst. Higher temperature and S/B ratio were observed to affect the amount of syngas. More amount of hydrogen with negligible CO_2 was observed by them because of utilization of adsorbent and catalyst. Several researchers have analysed sugarcane gasification process, but effects of more than two parameters are not found to be reported in the literature. It is observed from the literature [26] that researcher have considered silica sand as bed material and modeled biomass gasification for bubbling fluidized bed reactor where stoichiometric ratios i.e. Equivalence ratio, steam to biomass ratio and temperatures are considered in the range of 0.24 – 0.38, 0 – 0.63 and 700 - 840^{o}C respectively. According to them the gases released during devolatilization are found to affect the overall performance of the gasifier.

In some coconut-producing areas, away from the reach of a national power company, husk gasification of biomass feedstock has a high potential to be used as an energy source, as seen from the characteristics of coconut husks themselves [46]. . Lignin is the most powerful material in biomass. Lignin is highly resistant to degradation, biological, enzymatic, or chemical. Because of the relatively high carbon content compared to cellulose and hemicellulose, lignin has high energy content [10]. Therefore it is thought of analyzing the energy content of coconut coir through proper technology i.e. Gasification which show potential for use as fuel in power plants. Tooy et al.[47] gasified coconut husk to produce gas using a downdraft gasifier which was further used to generate electricity. Reactor temperature, tar volume produced, bioreactor gas produced, gasification performance, and efficiency in energy production were studied which revealed a 62% reduction in diesel fuel consumption. The diesel fuel machine generated by gas had a capacity of 10 kW. Dhurai et al.[48] carried out biomass gasification of coconut shell to produce product gas which was further used for combustion in a burner system. Feasibility of coconut coir dust as feedstock for 'Entrained Flow Gasification System' was studied by D. Singh [49] where the effect of equivalence ratio on syngas composition, adiabatic flame temperature, calorific value and

rate of gas generation was analysed. High quality results indicated the suitability of the process for industrialization. Hydrogen production potential from agricultural wastes (coconut coir, palm kernel shell) was analysed by air gasification technology using a fluidized bed gasifier [13].

2.8.2 Computational Investigations

The flow and reaction patterns in an entrained flow biomass gasifier have been simulatedby Fletcher et al. [10] based on the CFX package where the phenomena of turbulent fluid flow, heat transfer, species transport, devolatilization, particle combustion, and gas phase chemical reactions are described. Biomass particulate is modelled via a Lagrangian approach. Detailed information on the gas composition and temperature at the outlet are also obtained from this model. The inert sand bed is modelled as a static isotropic porous media containing prescribed spherical volumes to model the presence of rising bubbles in a bubbling fluidized bed. The emulsion phase of the fluidized region was modeled as a porous region, while stationary ”free” areas were prescribed to account for the rising gas pockets (bubbles) inside the bed [50]. The biomass particles are modelled as Lagrangian particles. The drying and devolatilization of biomass, heterogeneous reactions of char and a single reaction in the gas phase converting water and methane into carbon monoxide and hydrogen are taken into account by this model. The simulated exhaust gas concentrations for a 3D gasifier are found to agree reasonably well with the measured data for H_2, O_2, CO_2, and H_2O.

Taghipour et al. [51] studied the bed hydrodynamics of a two-dimensional gas–solid fluidized bed both experimentally and computationally. A multi-fluid Eulerian model incorporating the kinetic theory for solid particles was applied to simulate the gas–solidflow. Syamlal–O’Brien, Gidaspow and Wen–Yu drag functions were used to calculate the momentum exchange coefficients. The solid-phase kinetic energy fluctuation was characterized by varying the restitution coefficient values within 0.9 to 0.99.A CFD model for fluidized bed biomass gasifier is developed and the simulations were carried out to obtain the optimal condition for production of hydrogen rich gas [52] where the effects of the steam to biomass ratio, the equivalence ratio and the size of the biomass particles on the hydrogen yield were studied. The distributions of the hydrogen inside the gasifier were also described at different conditions.

An overview of different CFD studies on thermo chemical biomass conversion including gasification and combustion processes in, e.g., fixed beds, furnaces, fluidized beds and wood stoves was carried out [53]. Most of the cited work use commercial CFD codes with Euler–Lagrange modeling approaches. It was concluded that CFD can be used as a powerful tool to

predict biomass thermochemical processes as well as to design thermochemical reactors. CFD has played an active part in system design including analysis the distribution of products, flow, temperature, ash deposit and NOx emission. The CFD model results were observed to be satisfactory and in good agreements with the experimental data in most of the cases. Hamzehei et al. [54] applied CFD with a multi fluid Eulerian modeling incorporating the kinetic theory of solid particles to simulate the unsteady state behavior of two dimensional non-reactive gas–solid fluidized bed reactor. Syamlal-O'Brien drag functions were used to calculate momentum exchange coefficients and finite volume method was applied for discretization. The effects of particle size and superficial gas velocity on bed dynamics were also studied. The gasification of wood in a fluidized bed gasifier was modeled with the Eulerian multiphase approach [55]. The dispersed solid phase within the reactor was modeled as three continuous phases, i.e., one phase representing wood and two char phases with different diameters. Comparison of 2D simulation results with experimental data was carried out for the influence of initial bed height, wood feeding rate, and reactor throughput on product gas concentrations and temperature. Different model parameters like thermal boundary conditions, primary pyrolysis kinetics, and secondary pyrolysis model were also considered. Euler–Euler multiphase CFD model is proposed for continuous fast pyrolysis of biomass in a fluidized-bed reactor to determine the yields of tar, gas, and char [56 -58].A lumped, multi-component, multi-stage kinetic model is applied by Xue et al. [56] to describe the pyrolysis of a biomass particle while Boateng et al. [57] predicted the spontaneous emergence of pyrolysis vapors, char and non-condensable (permanent) gases and confirmed that the bio-oil vapor evolution is accomplished in a few seconds which occupies two-thirds of the spatial volume of the reactor as widely reported in the open literature.

A combined Eulerian/Lagrangian simulation method was used to compare against the experimental results [58] where effects of different process parameters were analysed and the overall thermochemical degradation process of biomass was observed to be influenced by local flow and particle properties. Thus this complex process could be analysed properly by Bruchmuller et al. [58] with the help of CFD modeling where superficial fluidization velocity was observed to be more important than the particle moisture content for the final bio-oil yield. The simulation of a bubbling fluidized bed reactor was carried out using three different meshing approaches and the results have been validated against the experimental measurements [59].An Eulerian–Lagrangian CFD model was described and validated against experimental findings. This model simulated the fluid dynamic behavior of gas and solid particles inside the freeboard region of a fluidized bed.The steam gasification of biomass was integrated with the hot gas cleaning systemin the reactor vessel [60].Fluidization tests with a cold model consisting of a bed of sand particles containing also a fraction of Geldart group A/C fine powder were carried out batch wise in a 10-cm ID column, at different operating conditions,

to quantify solid ejection into the freeboard zone. A Particle Elutriation Model (PEM) was proposed to set the boundary condition at the bed surface (freeboard inlet) for fine particles. A k-ε model was used for turbulence. A Discrete Random Walk (DRW) model was implemented to consider the particles turbulent dispersion. 2D simulations of the freeboard were carried out at different static bed height(10 and 20 cm), superficial gas velocity (11 and 16 cm/s) and mass of fines initially charged inside the bed.

The Eulerian–Lagrangian approach in CFD modeling is employed to study the gasification performance in a lab-scale pine gasifier. A series of simulations have been performed with some critical parameters including temperature, equivalence ratio and steam to biomass ratio. The model predicts product gas composition and carbon conversion efficiency in good agreement with experimental data [61]. Guodonget al. [62] addressed a multi scale coupling framework for modeling of large-size biomass particle gasification in fluidized beds, where the large diffusions due to the temperature and concentration gradients in large-size particles strongly affect gasification process. Sand particles were used as thermal carrier to keep the surrounding temperature statistically stable. The temperature was observed to be the dominating factor and the surrounding gas velocity was a secondary one to the gasification.
A 2D computational fluid dynamics (CFD) model was developed to study the gasification processin a downdraft configuration, considering drying, pyrolysis, combustion, and gasification reactions [63].An external heat source and superheated air combined with steam were used by them for the high-temperature agent gasification which resulted in a limited combustion need in the gasifier and produced syngas with a high H2 fraction and low tar content. Singh et al. [64] carried out an extensive review of CFD modelling to study combustion and gasification in fluidized beds. Janajreh et al. [65] used wood chips in a small scale, air blown, downdraft gasification system to investigate the conversion efficiency. High fidelity numerical simulation using CFD was used to model the Lagrangian particle coupled evolution. The temperature distribution and the evolution of species are computed and compared with the experimental results and with the ideal equilibrium, zero dimensional case. The role of CaO as a CO_2 sorbent along with other operating parameters, temperature (T)and steam to biomass (S/B) ratio on yield of hydrogen/tar production in a bubbling fluidized bed steam – biomass gasification was studied [66]. Compared to a bed of sand alone, a 20% higher H_2 concentration, an almost double H_2 yield and a 67% reduction in tar content were obtained when a bed of CaO was used. Moreover, it was observed that shifting the tar species from higher to fewer ring structures as a result of in-bed CaO could reduce tar dew point by 11 ^{0}C and tar carcinogenic potential by almost 60%.Loha et al. [67, 68, 69, and 70] carried out CFD modeling for biomass samples using a fluidized bed gasifier. These researchers have analysed different aspects such as (i)hydrodynamics of gas–solid flow in a

bubbling fluidized bed with Geldert B particles by applying the kinetic theory of granular flow (KTGF), (ii) the gasifier's behaviour including fluidization, thermal and chemical characteristics, (iii) the effect of the elasticity of particle collision on the hydrodynamic behavior of a bubbling fluidized bed with different restitution coefficients and (iv) comparative investigation of different modelling approaches applied to the fluidized bed gasification systems viz. the equilibrium model and the rate based or kinetic model.

The physical and chemical processes of particle gasification and their interaction with the reactive gas flow are modeled within a multi-fluid framework derived from kinetic theory of granular flows for biomass gasification using a fluidized-bed gasifier [71]. The effects of different parameters viz. air/biomass mass flow ratio, reactor temperature, and biomass moisture content are simulated and analyzed by these researchers so as to study their influence on gas composition and product yields at the gasifier outlet. Couto et al. [72] carried out CFD simulation to investigate the influence of oxygen-enriched air on a biomass gasification process. Hydrogen and nitrogen molar fractions were observed to decrease as a function of the oxygen content and carbon dioxide showed the opposite trend. On the other hand, there was only a slight increase of the methane molar fraction. Again the cold gas efficiency was observed to increase with the oxygen content and decrease slightly with the steam to biomass ratio. Ismail et al. [73] investigated the performance of an updraft gasifier experimentally and numerically using the kinetic theory of granular flow (KTGF) and k-ε model. They observed that the developed model is a promising tool for simulating the biomass gasification/combustion within the gasifier. Gas-solid interaction along with thermal flow behavior and gasification reactions are investigated for rice husk using a fluidized bed gasifier [74] where bed expansion, velocity profile, solid movement, temperature profile and species mass fractions are focused.

2.9 Objectives of the present work

On reviewing the various literatures on biomass gasification, it is thought of analyzing effects of several possible parameters on the yield of syngas produced during gasification using different biomass samples viz. rice husk, rice straw, saw dust, wood chips, sugarcane bagasse and coconut coir. In any process cost comes into the picture at the end. If the process is not cost effective, nobody will be interested to commercialise the production process. Without prior information regarding the fuel value of the biomass sample and energy output from the gasification technology it will lead to wastage of time and money. That is why it is essential to analyze the energy content of the sample and syngas to be produced through gasification before selecting any feed sample for gasification. Hence it is required to calculate the net energy output from the selected process and thereby to make a cost analysis.

Therefore, it is planned to analyze the composition of product gas obtained through gasification for different samples under different operating conditions

That is why energy balance and energy audit are planned to be carried out along with the biomass gasification process in the present work. Air requirement for gasification process is dependent on the C- H- O contents of biomass sample for which it is essential to determine the equivalence ratio.

Equivalence ratio (ER) is defined as the ratio of actual air flow to the gasifier and theoretical or stoichiometric air required for complete combustion of the biomass sample.

Thus, it is essential to have information on C-H-O contents of biomass feed sample for carrying out accurate energy balance calculations. In other words it can be said that stoichiometric amount of air needed for the combustion of the biomass sample can be determined if C-H-O contents or chemical formula for the biomass sample is known [75]. To check the proficiency of the specially designed in-house FBG various efficiencies are required to be calculated. To analyse the gasifier performance in terms of different efficiencies and to check the economic feasibility of the gasification process for various operating conditions it is planned to have an in-house gasifier instead of getting a readymade one. In-house equipment provides scopes for varied conditions. Therefore a fluidized bed gasifier is planned to design as per the requirement. Thus the objectives are summarized as follows.

2.9.1 Summary of Objectives

i. To design and have installed a fluidized bed gasifier for varied operating conditions.
ii. To carry out proximate/ultimate analysis and to determine the chemical formula of different biomass samples.
iii. To carry out gasification experiments using all the selected biomass samples in the FBG.
iv. To study the effect of different system parameters on the yield of syngas.
v. To carry out CFD simulation for hydrodynamic behaviour and thermal flow behavior of the biomass samples in a Fluidized Bed Gasifier.
vi. To carry out mass balance and energy balance calculations for all the biomass samples
vii. To carry out energy report for the gasification process and to determine the gasifier performance.
viii. To validate the gasifier design against experimental and CFD modeling results

Chapter -III

Design of Fluidized Bed Gasifier

A Fluidized Bed Gasifier is designed and fabricated for carrying out experiments on biomass gasification. The design is carried out according to information available in the literature [76] with some innovative modifications. The preliminary operating conditions such as fluidization velocity and equivalence ratiorequired for the proper gasification process on pilot scale are also considered. Minimum fluidization velocity and terminal velocity of the particles are calculated as per the literature [77]. The maximum value of the superficial velocity of the gas is determined as per expression given by Souza – Santos [78]. The superficial velocity of the gas to be used for the gasification operation is established considering the maximum and minimum expanded bed heights of the fluidized bed [76].

3.1 Design parameters

Minimum fluidization velocity, terminal velocity of particles, fluidization velocity and bed height are key parameters for proper design of a fluidized bed gasifier. The aim of the design is to have a bubbling fluidized bed. These parameters are discussed below.

3.1.1Minimum Fluidization Velocity

The lower limit of the superficial velocity of the gas that flows through the particle bed is calculated separately for the bed material and the biomass sample using the following expression [77].

$$U_{mf} = \frac{d_p^2(\rho_p - \rho_f)g \times \epsilon^3 \times \varphi^2}{150.\mu(1-\varepsilon)} \quad (3.1)$$

3.1.2 Terminal Velocity of the Particles

The maximum value of the superficial velocity of the gas is determined for both the materials of the bed depending on the Reynold number of the particle (for 0.4 < Re < 50) [78].

$$U_t = d_p \left[\frac{4(\rho_p - \rho_f)^2.g^2}{225.\rho_f.\mu}\right]^{1/3} \quad (3.2)$$

3.1.3Fluidization Velocity during the Gasification

The superficial velocity of the gas to be used during the gasification operation is established considering the relation between the expanded and minimum heights of the fluidized bed [76].

$$\frac{H}{H_{mf}} = 1 + \frac{10.978\left(U_f - U_{mf}\right)^{0.738}.\rho_p{}^{0.376}.d_p{}^{1.006}}{U_{mf}{}^{0.937}.\rho_f{}^{0.126}} \quad (3.3)$$

For the bubbling fluidized bed, a restriction as mentioned below is suggested in the following expression [77]. That means the ratio of bed height under any condition cannot exceed the limits of 1.2 to 1.4 times the bed height under minimum fluidization.

$$1.2 < \frac{H}{H_{mf}} < 1.4 \quad (3.4)$$

Initially, a value of 1.25 is selected for the above ratio and the equation (3.3) is solved to determine the value of U_f. The fluidization velocity is finally considered as 0.7 m/s for biomass samples.

3.1.4 Overall Height of the Reaction Chamber

The overall height of the reactor is determined by the following expression [77].

$$H_t = TDH + H \quad (3.5)$$

The maximum expanded height of the bed is assumed to be 0.5 m, being twice the internal diameter of reactor, with the purpose of diminishing the slugging phenomena.

3.2 Outlet Dust Separation

The fluidized bed gasifier requires a high efficiency cyclone which can collect the particulate materials that could be released during the gasification process. Cyclone separator is designed as per the literature [79]. A cyclone with the geometric dimensions is presented below in the Table-3.1 and 3.2.

Table-3.1 Assumed parameters for cyclone separator design

Parameter	Value
Gas inlet velocity (m/s)	15 - 30
Pressure drop (kPa)	< 2.5
Collection efficiency (%)	> 90

Table-3.2 Design data (Dimensions) of cyclone separator

Parameter	Value
Cyclone diameter (cm)	20
Cyclone gas exit diameter (cm)	9.5
Cyclone body cylindrical height (cm)	36
Cyclone total height (cm)	69
Cyclone solids exit diameter (cm)	3
Separation efficiency (%)	99.7
Pressure drop (kPa)	0.45

From the mass balance of the product gas in the gasification process, the density and the volumetric flow rate of inlet gas to the cyclone were calculated at the operating conditions (approximately 750ºC and 101,325 kPa). Table - 3.2 shows the dimensions of the designed cyclone, along with its efficiency and pressure drop. A cyclone separator with the dimensions is shown in Fig. 3.1.

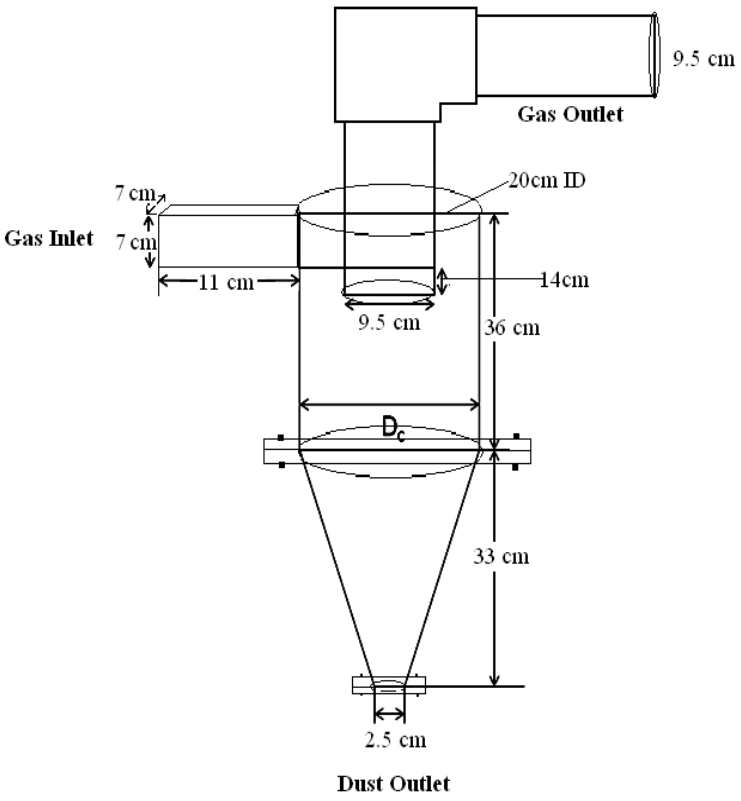


Fig.3.1 Design of Cyclone Separator

3.3 Biomass Feeding System

The feeding system for the biomass as shown in Fig. 3.2 consists of a hopper, a storage chamber and a feeding assembly having a gear box and a feeding screw. The screw is driven by a motor with a variable frequency drive (VFD) as a speed controller. The feeding screw introduces the biomass and the bed material to the chamber and operates like a dosingscrew at a greater speed to avoid feed accumulation which causes system blockages [80 & 81]. Schematic diagram is given in Fig.3.2.

3.3.1 Screw Sizing:

The relation between the biomass and bed material flow with the bed diameter, pitch, fillet height and revolutions of the screw has been described by the following expression [82].

$$\dot{m}_{bm} = 60.\pi.s.n.\phi.\rho_{bm}.(D.h - h^2) \quad (3.6)$$

Where,

D	screw outlet diameter, cm
U_{mf}	Minimum fluidization velocity, cm/sec
g	Acceleration due to gravity, m/sec^2
φ	Sphericity
μ	Air viscosity (approximately 750^{0}c and 1atm), kg/m-sec
ρ_{bm}	Biomass density, kg/cm^3
ρ_p	Particle density, kg/cm^3
ρ_f	Fluid density, kg/cm^3
ε	Porosity
h	Fillet height in cm
s	step screw, m

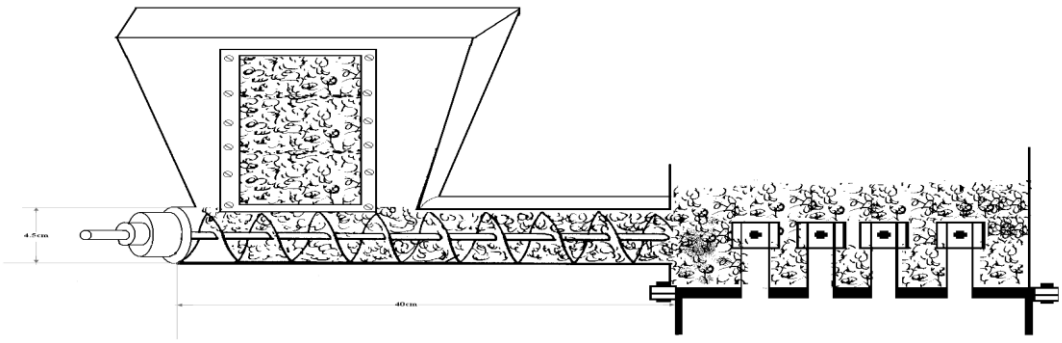


Fig.3.2 Design of Screw Feeder

3.4 Air Distribution (Bubble Caps)

A bubble cap type distributor is selected. Orifices are of 5mm diameter openings. There are seven orifices on the riser through which air passes and then air is distributed uniformly into the reactor by creating eddy mixing of biomass and bed material. Six numbers of bubble caps are arranged in a circular pattern towards the outer periphery of the distributor making the boundary with seventh one at the center (Fig. 3.4). This alternative was selected due to its convenience for use with high temperatures and the advantage of preventing the backflow of bed material towards the plenum The data for the distributorplate design as obtained by using the model calculation proposed in literature[1] are listed in Table - 3.3 and Table -3.4. Design diagram of the distributor and bubble cap are shown in Fig. 3.3.

Table- 3.3 : Design parameters for the air distributor.

Parameter	Value
Fluidization velocity ($m.s^{-1}$)	0.5
Minimum fluidization velocity ($m.s^{-1}$)	0.05
Minimum fluidization height (m)	0.35
Particle density ($kg.m^{-3}$)	120 - 2,650
Mean particle size (μm)	200 - 10000
Bed porosity	0.46
Bed zone diameter (m)	0.15

Table- 3.4 : Calculated parameters for the distributor plate.

Parameter	Value
Pressure drop in the bed (KPa)	5.5
Orifice diameter (mm)	5
Number of orifices openings on pipe	4
Number of bubble cap pipes	7

3.5 Cold Model Gasifier

The bench scale fluidized bed gasifier (both cold and hot model) with nominal capacity of 20 kW has been designed. Using the design data, different parts the cold model fluidized bed gasifier were fabricated, assembled and installed in the laboratory. The schematic diagram of the cold model gasifier is shown in Fig.3.4. A cold model gasifier was designed and fabricated first using Perspex as material of construction. The effect of different system parameters on the bed hydrodynamics were studied in the same unit. The bed behaviours were also visualized through transparent wall of the column. Finally optimum conditions were found out by experimental and computational analysis.

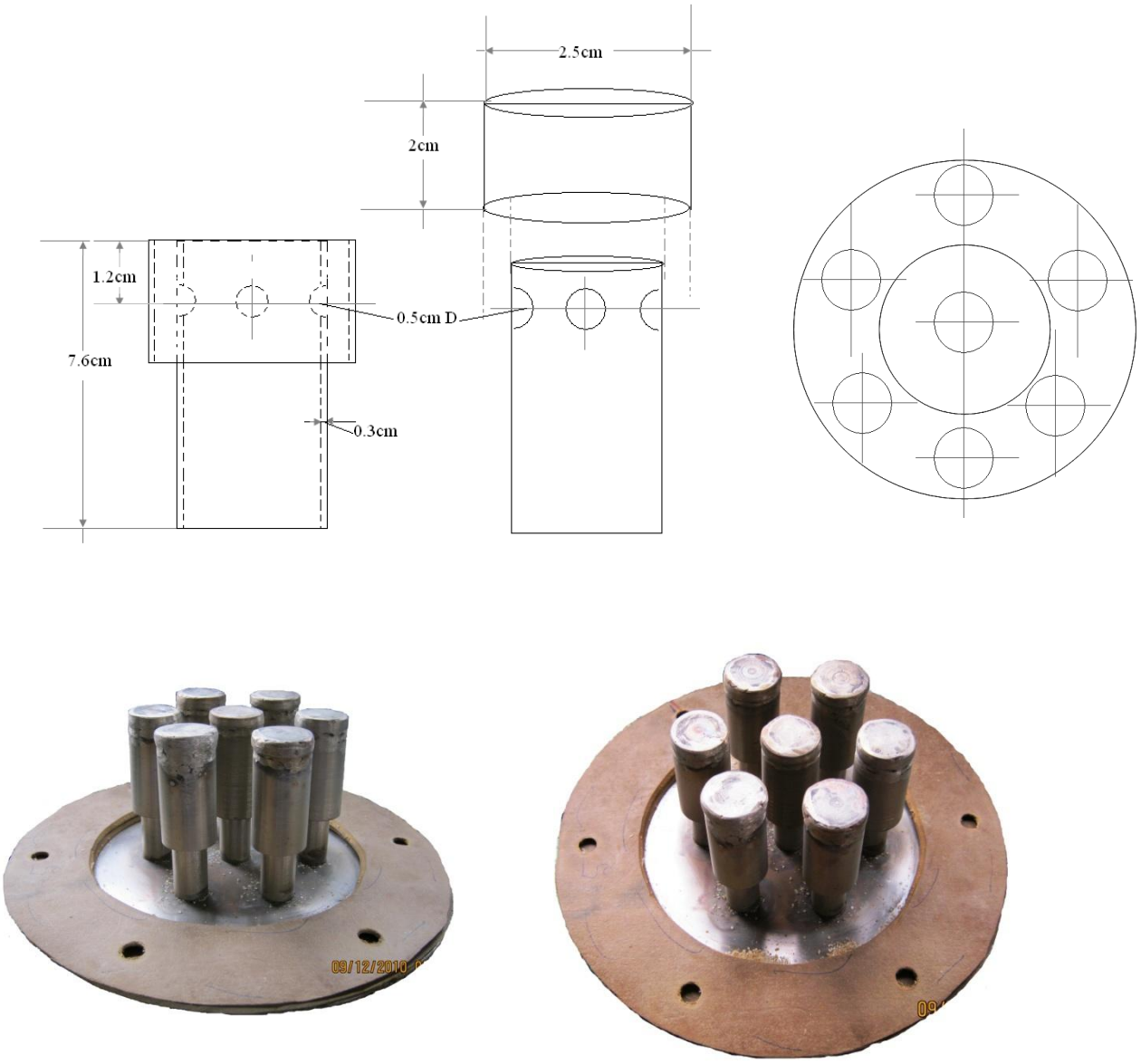


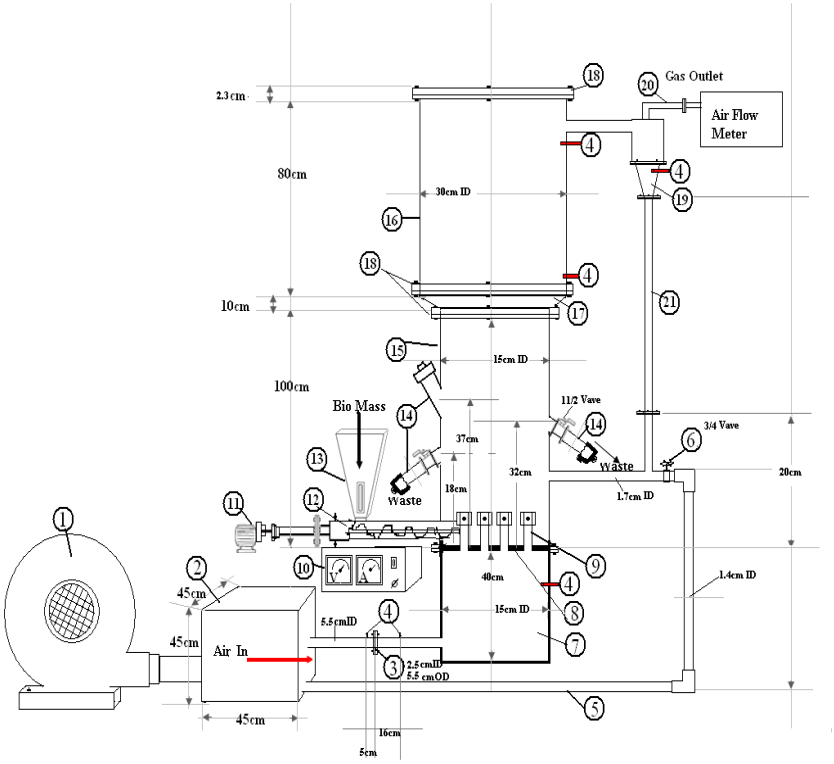
Fig.3.3 Design of Bubble Cap and Distributor Arrangement

3.6 Hot Model Gasifier

The hot model gasifier (real one) was then fabricated with the same design data in addition to a steam generator of 5 liter capacity. Then the hot model gasifier was fabricated from mild steel material with 7cm- inside lining of high alumina sand for carrying out several experiments. The inside dimensions were kept same as the cold model gasifier. Heat and steam supply arrangements were additionally provided. Experiments were then carried out

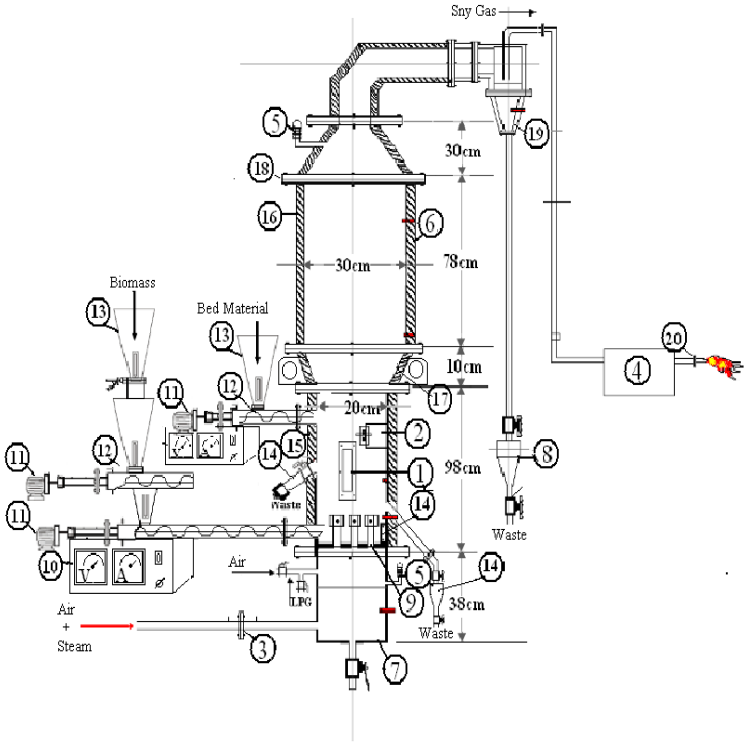
with the established optimum variables using biomass sample. Finally production of Hydrogen was ensured with several testing.

Photographs of the laboratory units for both cold and hot model are shown in Fig.3.6 and Fig. 3.7 respectively.



Sl. No.	Description
1.	Blower
2.	Header
3.	M.S. Flange with Orifice Meter
4.	Pressure tapering for Manometer
5.	Air bypass
6.	Air by pass controlling valve
7.	Air mixing chamber
8.	Distributer Holding flange
9.	Bubble Cape
10.	Motor Controller with Voltammeter and Ammeter
11.	Motor
12.	Screw Feeder
13.	Hopper
14.	Drain System
15.	Bed Zone
16.	Free Board Zone
17.	Reducer
18.	Flange Joints
19.	Cyclone Separator
20.	Gas Outlet
21.	Dust Recycle System

Fig.3.4 Design of Cold Model Fluidized Bed Gasifier



Sl. No.	Description
1.	Site Glass
2.	Firing gate
3.	M.S. Flange with Orifice Meter
4.	Gas Analyzer
5.	Safty Valve
6.	Insulation
7.	Air mixing chamber
8.	Cooling storage before drain
9.	Bubble Cape
10.	Motor Controller with Voltammeter and Ammeter
11.	Motor
12.	Screw Feeder
13.	Hopper
14.	Drain System
15.	Bed Zone
16.	Free Board Zone
17.	Reducer
18.	Flange Joints
19.	Cyclone Separator
20.	Gas Outlet
21.	Dust Recycle System

Fig.3.5 Design of hot model fluidized bed gasifier

Fig.3.6 Cold model fluidized bed gasifier (Laboratory Unit)

Fig.3.7 Hot model fluidized bed gasifier (Laboratory Unit)

Chapter-IV

Experimental Aspects

Biomass gasification is carried out in the in-house gasifier which is an indigenous one to determine the yield of synthesis gas and its composition. Various experiments are conducted using different biomass samples and bed materials in the gasifier. Different types of commonly available biomass samples are collected from the local area. These samples are characterized and pretreated before gasification process to estimate the amount of energy available in the biomass sample. Proximate and ultimate analyses of the biomass sample are most important to know the percentage of basic elements present in the samples. Sizing of feed sample is essential depending upon the feed port size. In addition, drying is required to keep the moisture within the permissible limit to make the gasification process possible. Some experiments are already conducted with this indigenous gasifier in the laboratory for trial run and compared with Aspen plus Simulations [83].

4.1 Materials

The following feed materials and bed materials have been used in the fluidized bed gasifier for gasification experiments. Sample pictures of these materials are shown Fig.-4.1 and 4.2.

4.1.1 Bed Materials:

Sand, Dolomite, Red Mud and their mixtures in different proportions by weight are used as bed materials in the present work and are shown below in Fig. 4.1.

Fig.4.1: Sample pictures of different bed materials

4.1.2 Feed Materials (Biomass Samples):

Different biomass samples such as Sawdust, Rice Husk, Rice Straw, Wood Chips, Sugarcane Bagasse and Coconut Coir are handled for the gasification experiments and shown below in Fig. 4.2

Fig.4.2: Sample pictures of different feed (biomass) samples

4.1.3 Fluidization Media:

Steam and Air are used as the fluidization media. Rice husk and saw dust as available were used directly in the gasifier. But Rice straw, wood chips, sugarcane bagasse, coconut coir are sized to meet the required size by cutting.

4.2 Physical Properties

Physical properties like bulk density, mean particle size, sphericity and porosity of different feed samples and bed materials are measured before experimentation (Table-4.1). Ultimate analysis and proximate analysis are carried out for the characterization of the different biomass samples (Table-4.2 & 4.3). These analysis give information about the amount of hydrogen and fixed carbon content present in the biomass samples.

4.2.1 Ultimate analysis

Determination of total carbon, hydrogen, nitrogen, oxygen and sulfur percentages in the biomass sample is carried out by its ultimate analysis. The data obtained from ultimate analysis of the biomass samples are shown in Table - 4.2

4.2.2 Proximate analysis

Determination of moisture content, volatile matter, ash content and fixed carbon in the biomass sample is known as the proximate analysis. The proximate analysis results for different biomass samples are given in Table - 4.3. Composition and density of materials used as inert (e.g. sand) and catalysts (e.g. red mud) are given below (Table 4.4).

Table - 4.1 Physical Properties of Biomass and bed material

Property	Mean particle size (mm)	Apparent density (kg/m^3)	Porosity	Sphericity
Bed material				
Sand	0.38	2650	0.44	0.77
Dolomite	0.55	2800	0.36	0.79
Red mud	0.22	1290	0.42	0.72
Biomass				
Rice husk	0.53	426	0.81	0.37
Rice straw	5.0	153	0.46	0.56
Saw dust	0.81	244	0.7	0.45
Wood chips	5.0	481	0.47	0.1
Coconut coir	10.0	352	0.96	0.04
Sugarcane bagasse	10.0	120	0.62	0.01

Table-4.2 Ultimate Analysis of selected biomass samples

Biomass samples	C (%)	H (%)	N (%)	S (%)	O_2 (%)
Rice husk	38.45	4.96	0.82	0.18	55.59
Rice straw	38.6	4.55	0.47	0.21	56.17
Saw dust	45.78	5.32	0.16	0.07	48.65
Wood chips	46.23	5.7	0.22	0.12	45.2
Sugarcane bagasse	44.60	6.2	0.20	0.50	46.84
Coconut coir	43.76	5.8	0.40	0.22	47.12

Table - 4.3 Proximate Analysis of selected biomass samples

Biomass samples	Moisture content (%)	Volatile matter (%)	Ash content (%)	Fixed carbon (%)
Rice husk	7.34	56.37	15.83	20.46
Rice straw	9.38	69.53	3.04	18.05
Saw dust	8.8	87.57	1.94	16.45
Wood chips	8	74.34	1.8	16.8
Sugarcane bagasse	5	73.8	1.66	19.54
Coconut coir	5.3	76.8	0.9	17.0

Table 4.4 : Composition of bed materials

SL.NO.	Bed Materials	Particle Size (dp), μm	Particle Density (ρs), kg/m3
1	Red Mud	77	1300
2	sand	380	2650

Sand :

Composition	**Weight %**
Carbonates and shale	-53.0
Feldspars	- 27 .0
Quartz	- l7.0

Redmud :

Composition	**Weight %**
Fe2O3	30-60
Al2O3	10-20
SiO2	3-50
Na2O	2-10
CaO	2-8
TiO2	25

4.3 Different Parts of the Experimental Setup

The laboratory unit of fluidized bed gasifier consists of different components as its accessories. These are mentioned below.

4.3.1 Biomass Feeding System

The feeding system as shown in Fig.-3.2 is used for the gasifier and handles biomass samples in the particle size range of 200 to 10000micron.

4.3.2 Air Distribution (Bubble Caps)

A bubble cap air distributor as shown in Fig. 3.3 is usedat the bottom of the gasifier.

4.3.3 Air Blower :

A blower with controlling valve is used for continuous air supply. A blower of 750-1000 WGP, 5hp Motor with 2840 RPM (Make:Cromton Greaves Ltd.) as shown in Fig.4.3 has been used in the experimental setup. A special square type header of 45cm length is provided for the pressure head safety.

10 H.P.Blower:

a) Motor = 10 H.P.
b) Blower casing M.S. sheet
c) Base frame M.S. structure,
d) Empeller M.S. sheet,
e) Direct motor couple
f) statter star delta,
g) main switch 63 amp
h) foundation M.S. structure
i) Blower capacity 800mm WGP to 1000 mm WGP.

4.3.4. Steam Generator:

A steam generator of 5-liter capacity is used to produce steam (Fig.-4.4) which is supplied to the gasifier along with the air.

Supply of steam generator, capacity 5 kg. Including.

a) Mounting trolley with wheel,
b) Water tank M.S.
c) Main chamber SS-304
d) Top cover SS-304
e) Bottom cover SS-304
f) Pressure gauge,
g) Limit switch
h) Water gauge glass with measuring scale,
i) Water heater, 1000W to 3000W
j) Water pump
k) Insulation chamber outside

l) Steam outlet ½" pipe SS-304
m) Main chamber mounting stand
n) Water outlet pipe ½" dia with valve
o) safety valve 2 nos.(two) close type

Fig.-4.3 : Air Blower

Fig.-4.4 : Steam Generator

Temperature Measurement by Thermocouple:

1. Both side high temperature glass fixing for see the inside of furnace.
2. One side air lock gate for firing system.
3. One side lock gate for bed material clean.
4. Thermocouple- capacity upto 1200^0 C= 5 nos.
5. Display- upto 1200^0 C= 5 nos.
6. Control panel = 1 no.

4.3.5 Cyclone Separator :

This gasification system uses of a high-efficiency cyclone separator with vessel dia of 20cm, exit diameter of 10 cm and height of 36cm as shown in Fig. 3.1to collect the particulate materials which are passeddue to incomplete combustion during the gasification process.

Arrangement for LPG supply and firing point are also made as shown in Fig. 4.5. Three outlet points are located at different heights of the gasifier. Location of the sampling point is shown in Fig.-4.6.

4.3.6 Fluidized Bed Gasifier:

Schematic diagram of the experimental setupis shown in Fig. 4.7 and photograph of the laboratory unit of gasification unit isshown in Fig.4.8.

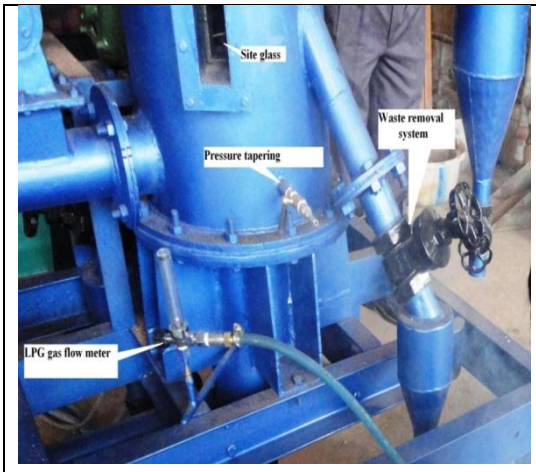


Fig.-4.5LPG injection point

Fig.-4.6 Sample point and sampling arrangement

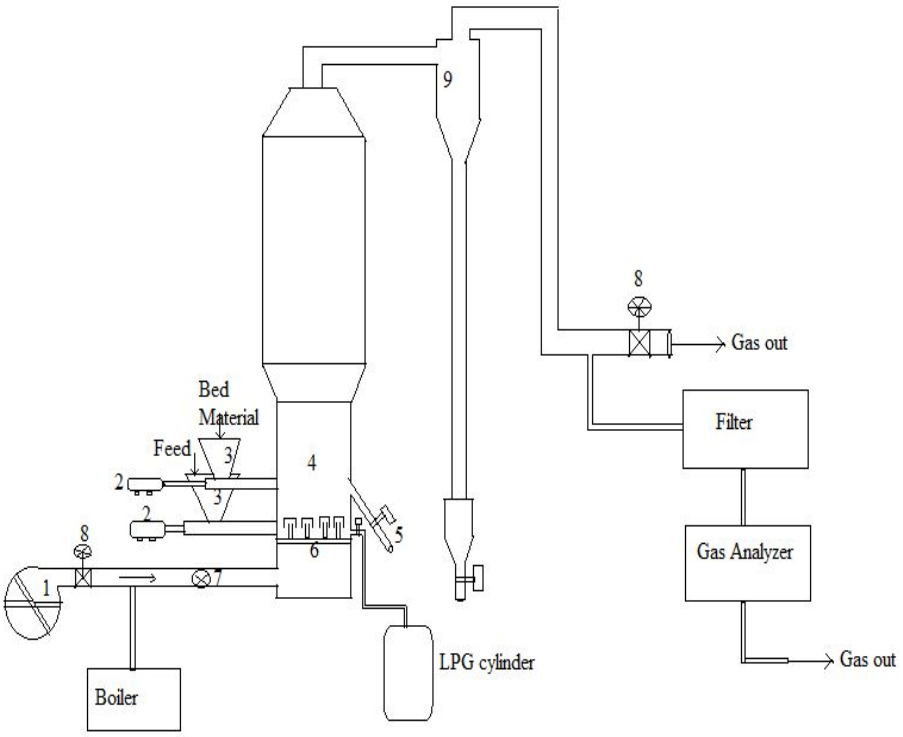


Fig. 4.7: Schematic diagram of the experimental setup for the gasification system.

1	Air blower
2	Motor
3	Screw Feeder
4	Fluidized bed gasifier
5	Continuous cleaning
6	Bubble cap
7	Orifice meter
8	Valve
9	Cyclone separator

Fig.4.8: Photograph of the experimental setup (Lab. unit gasifier)

4.4 Methods

The experiments are carried out in two levels,and CFD simulations are also performedin two levels. Experiments are conductedwith respect to the following aspects.

(a) Studies on the effects of parameters such as Steam to Biomass ratio (S/B ratio), equivalence ratio (ER) and temperature for different types of biomass samples viz.Sawdust, Rice Husk, Rice Straw, Wood chips, Sugarcane Bagasse and Coconut Coir.

(b) Studies on the catalytic effects on yield of syngas using different bed material (sand, thesand-dolomite mixture in different proportions and sand-red mud mixture in different proportions).

4.4.1 Operating Procedure:

Initially inert bed material is fed to the gasifier up to certain height (2.5 kg in the present work). Biomass sample is then fed continuously by the screw feeder. A specified quantity of water is addedto thesteam generator for generation of steam. Feedstocks in the gasifier are then ignited by LPG which is supplied at a flow rate of 10-12 LPH to preheat the bed material within the gasifier till the temperature reaches up to 550 – 600^{o}C. The

temperature of the bed material within gasifier increases gradually. By the time it reaches 800^{o}C steam starts pinching. When the temperature at the neck and outer wall of furnace reaches 900^{o}C gasifying agents are driven into the gasifier, and then the tests start up. The temperatures at seven different locations at different intervals of thetestare recorded. The gas yield is measured by a flow meter simultaneously. Usually, the steady state is reached at around 15minutes of startup and then gas sampling is done at 10minutes of thetime interval. Each operating condition is repeated twice to assure the quality of the experimental results and the average of these observed values are taken into consideration for further calculations. When the bed temperature increases to the desired level, the biomass feed samples are fed to the reactor by the feeder and the flow of liquefied petroleum gas into the gasifier is then stopped. The gasification process starts taking place. The gaseous product obtained from the cyclone separator is then passed through the gas analyzer to measure the composition of syngas,i.e., thepercentage of hydrogen, carbon monoxide, methane and carbon dioxide, etc., in the product gas. The operating parameters studied while experimentation are listed in Table-4.5.

Table No. 4.5: System parameters studied for gasification.

S.No.	Operating Parameter	Range
1	Temperature	500-800^{o}C
2	Bed material	3-4 kg.
3	Feed rate	10-15 kg/hr
4	Equivalence ratio	0.20-0.50
5	Steam to biomass ratio	0.5 – 2.5

4.4.2 Output Measurement:

The objective of the present investigation is to measure the yield i.e. hydrogen gas by varying different system parameters and thereby to determine the optimizing the gasification efficiency. Different parameters viz. the reaction temperature, bed materials for studying the catalytic effects, the percentage of stoichiometric air and steam supply in turn, the equivalence ratio and steam to biomass ratio are varied. Effects of these

parameters on the production of Syngas as well as the composition of Syngasare also observed during the experimentation. Picture of Gas Analyzer is shown in Fig.-4.9. Outputs obtained with different biomass samples are then analyzed for further calculations to know energy content of different biomass samples as well as the gasifier efficiency.

Gas cleaning system is nothing but gas conditioning device. It is an accessory for the gas analyser. Syngas is allowed to pass through the gas cleaning system where dust particles are separated out at first. As for the gas analyser gas Sample should be free from dust, water vapor and tar. The cleaned gas is then passed through the gas analyser. The used gas analyser is a portable biomass gas analyzer- Online and portable type model.

Model: Portable GAS3100P Syngas Analyser

Supplier:

ACE gas analysers Pvt.Ltd.

Opp. Bank of Baroda,

M.G. Road, Ghatkopar (W)

Mumbai-400086.

It measures the syngas composition on Nitrogen free basis as per the following.

O_2 % + CO% + CO_2% + CH_4 % + H_2 % + Gas Calorific value calculation in the one unit.

1. Measuring principle : CO/CO_2 /CH_4 /CnHm :NDIR,H_2 :TCD,O_2 : ECD
2. Measurement Range : CO:0-75%, CO_2 :0-40%, CH_4 :0-75%, H_2 :0-75%, O_2 :0-25%, CnHm:0-10%,
3. Precision: H_2 /CO/CO_2 /CH_4 /CnHm: 1%FS; O_2: 2%FS.
4. Resolution: O_2 /H_2 /CO/CO_2 /CH_4 /CnHm: 0.01%.

Gas analyzer system- It uses Electrochemical and NDIR principle to measure the gas concentration in ppm or %.

Fig.4.9 (A) : Gas cleaning system

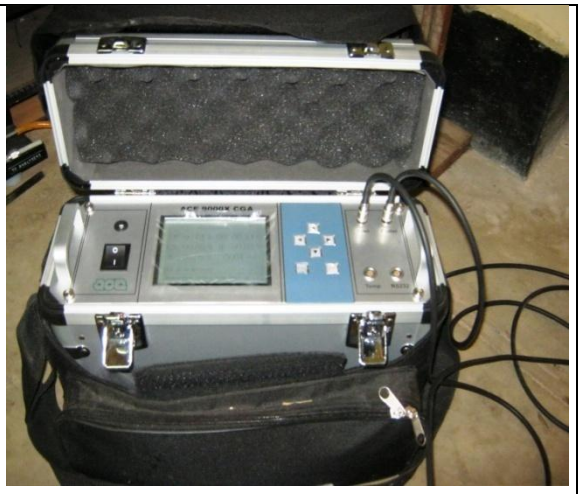


Fig.4.9 (B) : Gas analyzer system

4.5 Experimental Observations:

Experiments are carried out in two levels as described below (A) to study the effects of different parameters on different biomass samples, (B) to study the catalytic effects of different bed materials. The observed data is listed in Appendix-I.

4.5.1 With respect to Parameter effects on Syngas:

Temperatures observed in the different zones of the gasifier during the process are shown in Table-A-1 (Appendix-A).The temperature profile for the gasifier is shown in Fig. 4.10. The product gasses obtained from different feed materials within the gasifier are analyzed by using a Gas analyzer. As air is used as the fluidizing medium, the process is air gasification for which Nitrogen and Oxygen will be the major portion of the product gas. Therefore,nothing of all the observations and computation of Syngas composition are carried out on N_2 and O_2 free basis for different biomass samples. The variation of Syngas composition against temperature for these materials are shown in Fig.-4.11 and listed in Table-A-2(a-f) (Appendix-I). Effects of temperature on yield of different components for syn-gas have been compared for different feed materials in Fig.4.12. Effects of S/B ratio and equivalence ratio on different components of Syngas for different feed materials are also studied and listed in Table-A-3(a-f) and 4 (a-f) in Appendix-I) respectively. The respective plots are shown in Fig.4.13 and 4.14.

4.5.2 With respect to catalytic effects of bed materials:

In the previous section, the effects of different parameters on the yield of syngas from different biomass samples are studied with the sand as the bed material. Red mud and

dolomites are used along with the sand as the bed material to examine the catalytic effects on yield and composition of synthesis gas.A mixture of sand with red mud and dolomite in different proportions are considered in the present work. Experimental data for the composition of syngas against temperature with different bed materials for different biomass samples are observed and listed in Tables - B-1 to B- 6. These bed materials are observed to have catalytic effects on yield of syngas. These effects are compared in Fig.4.15 to 4.20 for different biomass samples.

4.5.2 Product Gas

Average flow rates of product gas for different biomass samples and their net heating values (NHV) are measured by using flowmeter and gas analyzer. These observations are listed in Table-4.6.

Table-4.6: Heating values and flow rates of product gas

Sl. No.	Biomass sample	HHV, MJ/kg feed	Avg. gas production rate, m^3/kg of feed	NHV, Kcal/m^3 of product gas
1	Rice husk	16.2	1.30	2365
2	Rice straw	16.78	1.28	2340
3	Saw dust	16.2	1.12	2586
4	Wood chips	15.6	1.15	2462
5	Sugarcane	20	1.4	2650
6	Coconut coir	19	1.45	2317

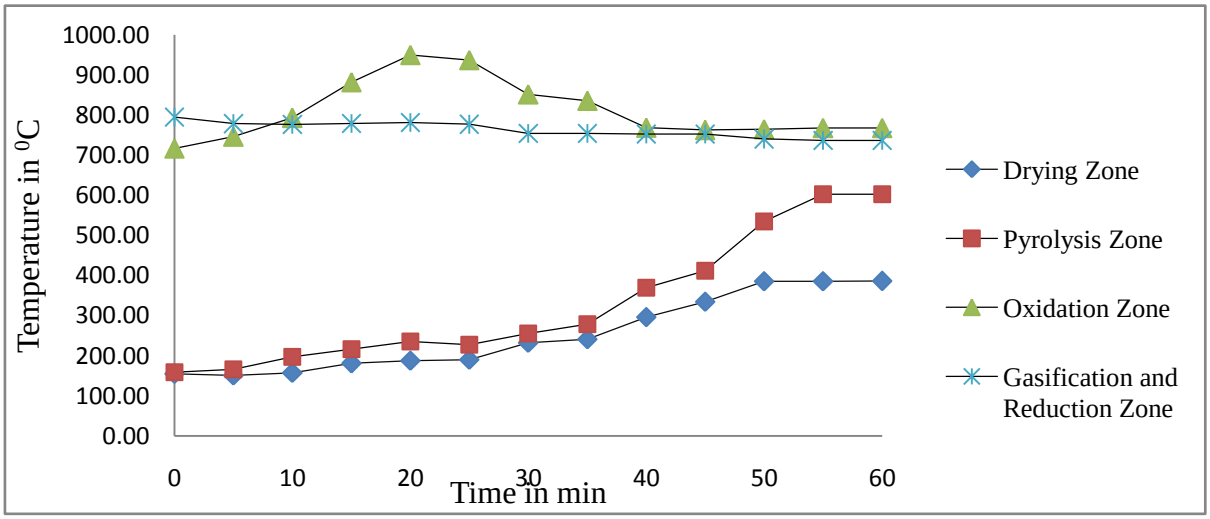
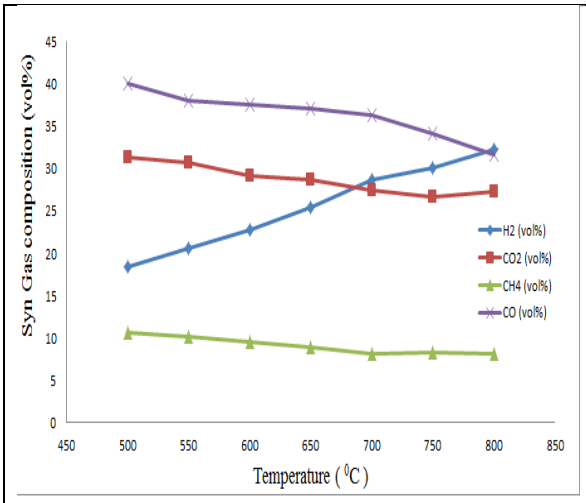
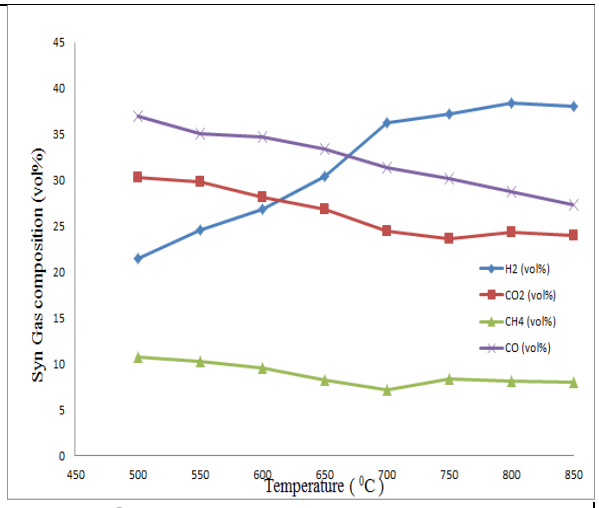


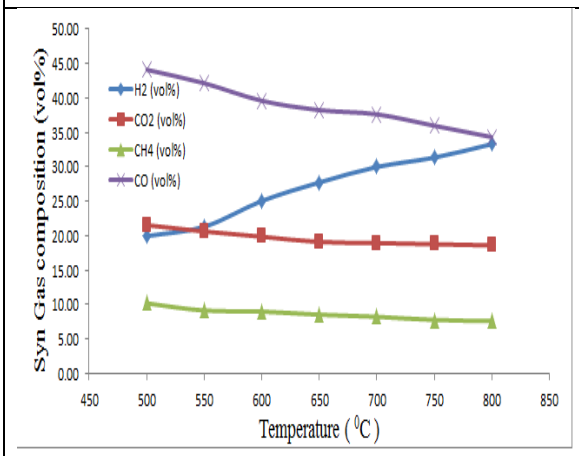
Fig.-4.10: Temperature profile for different zones within the gasifier



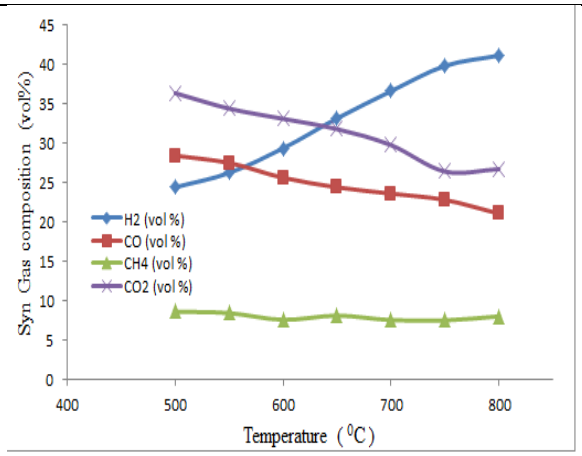
Rice Straw



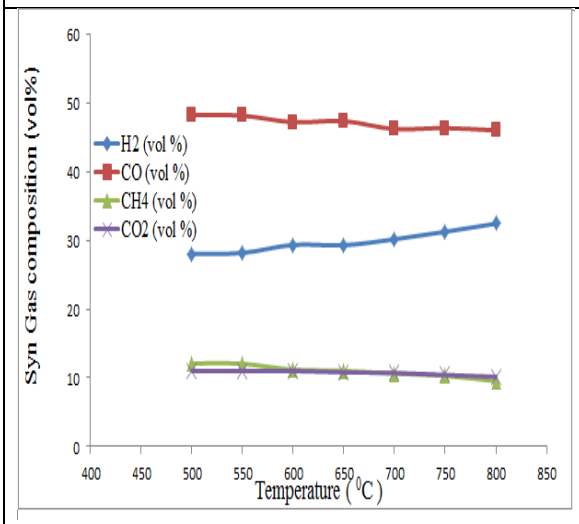
Rice Husk



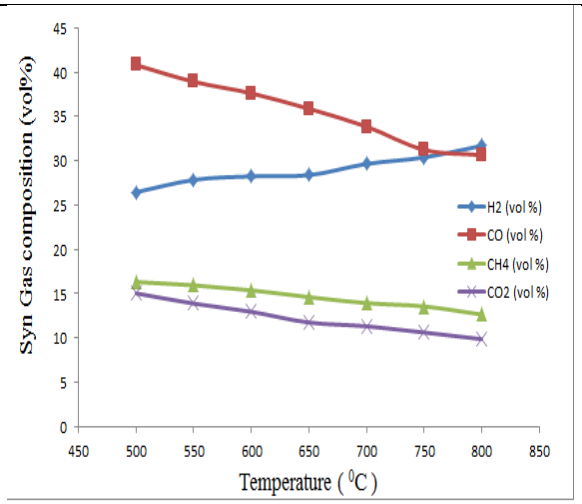
Saw Dust



Wood Chips



Sugarcane Bagasse



Coconut Coir

Fig.-4.11: Effect of temperature on different components of product gas for different biomass samples on Nitrogen and Oxygen free basis.

(A) Parameter Effects on Yield Of Syngas

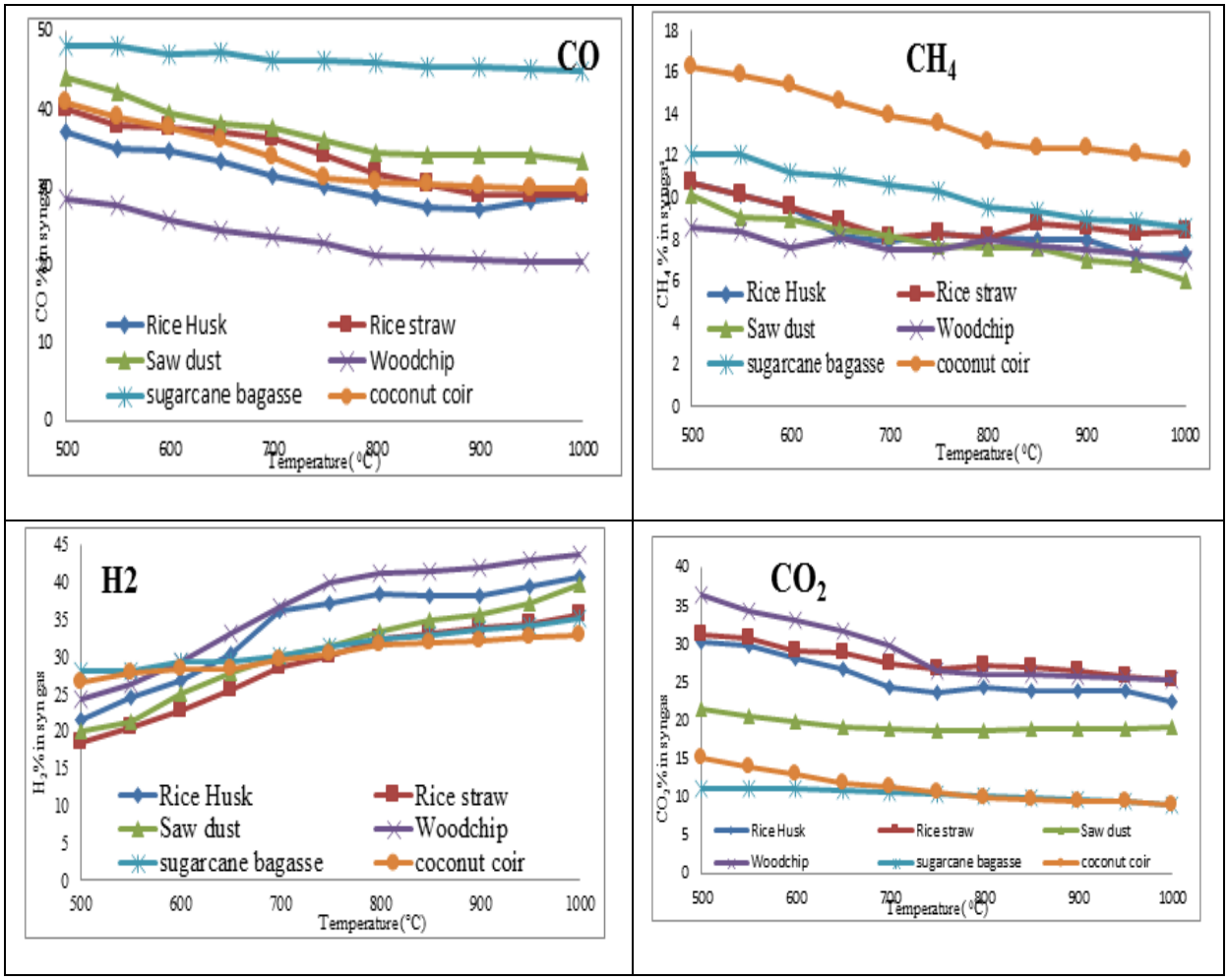
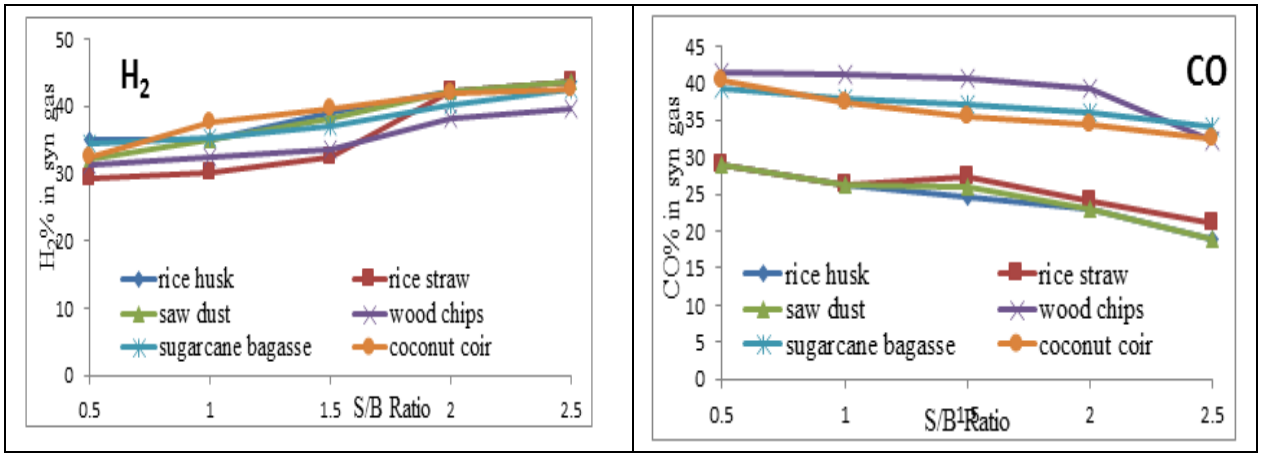


Fig.4.12 : Comparison of effects of temperature on yield of individual components for different feed samples at ER= 0.25, S/B ratio =0 and Feed Rate = 10kg/hr



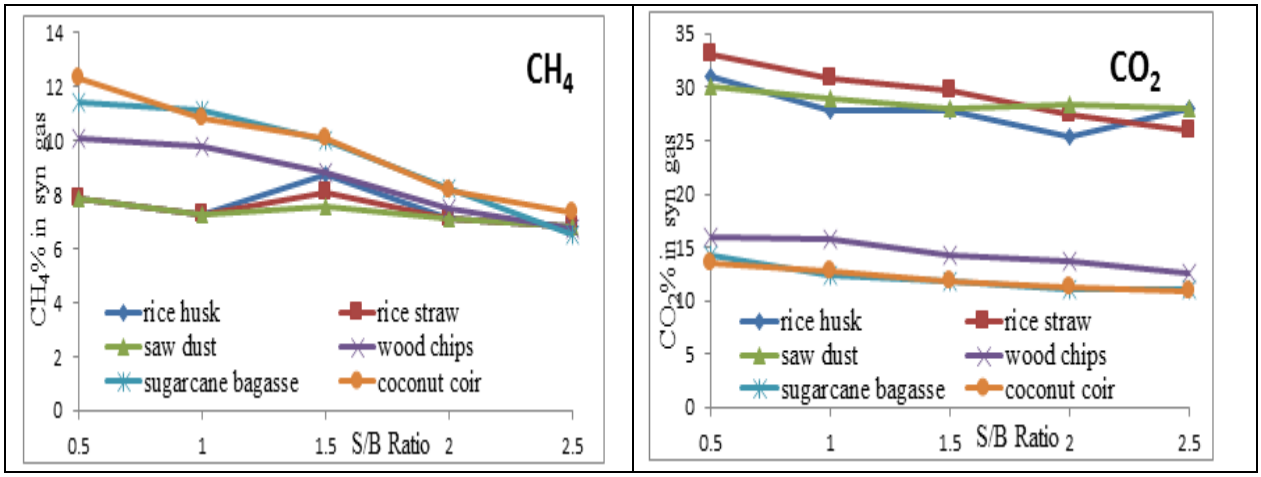


Fig.-4.13 : Comparison of effects of S/B ratio on yield of individual components for different feed samples at ER= 0.25 and Feed Rate = 10kg/hr

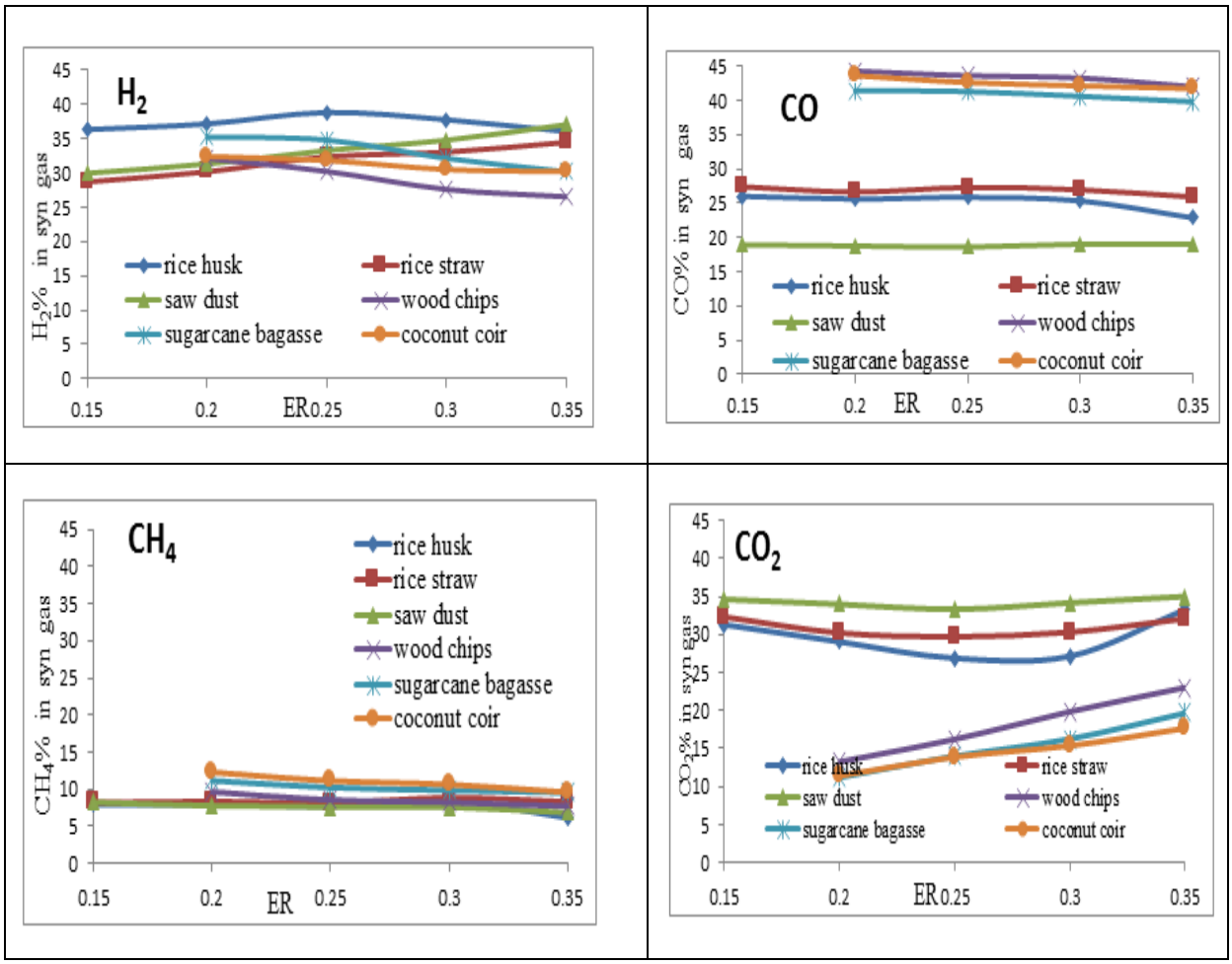
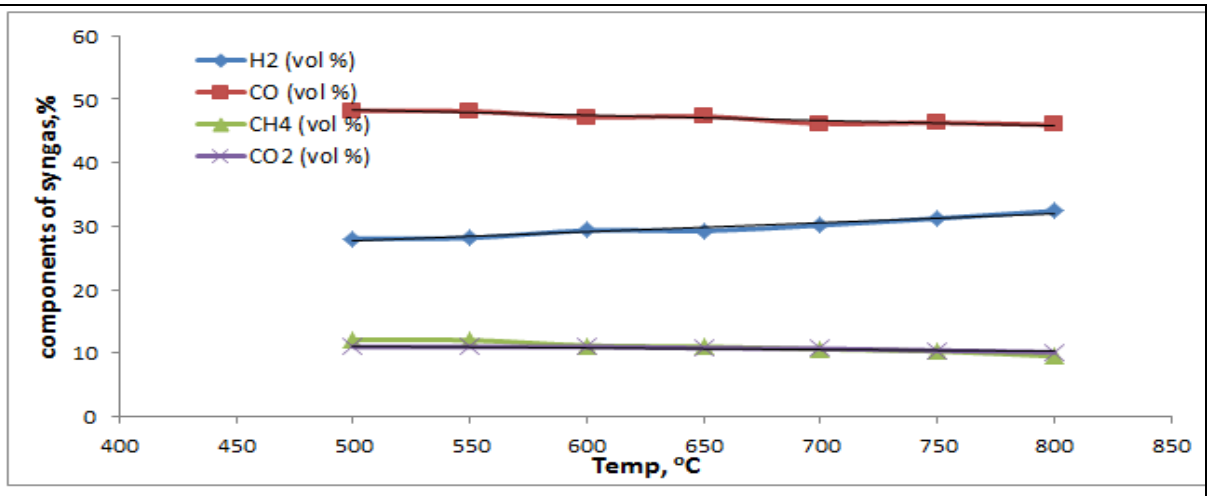
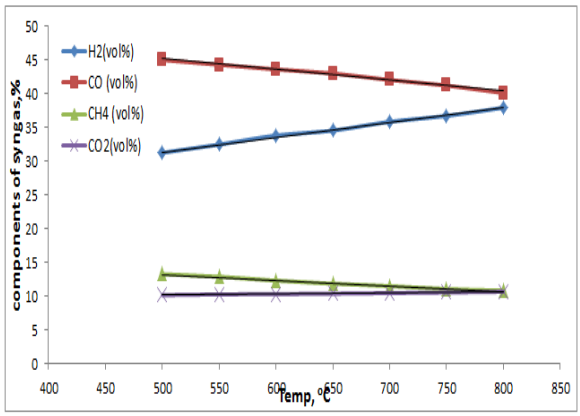


Fig.4.14 : Comparison of effects of ER on yield of individual components for different feed samples at S/B= 1.5 and Feed Rate = 10kg/hr

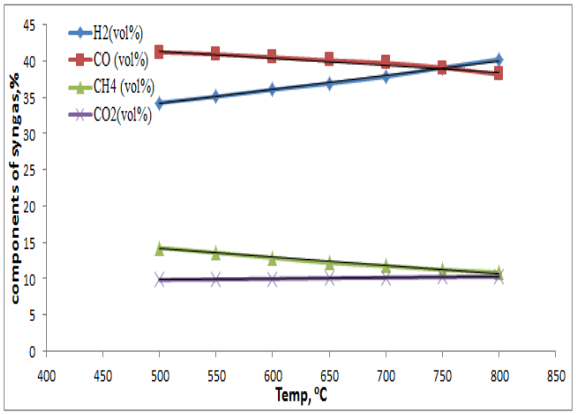
(B) For Catalytic Effects on Yield of Syngas



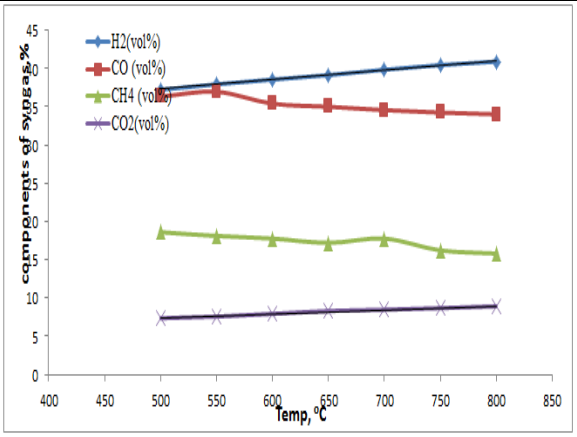
For Sand as bed material



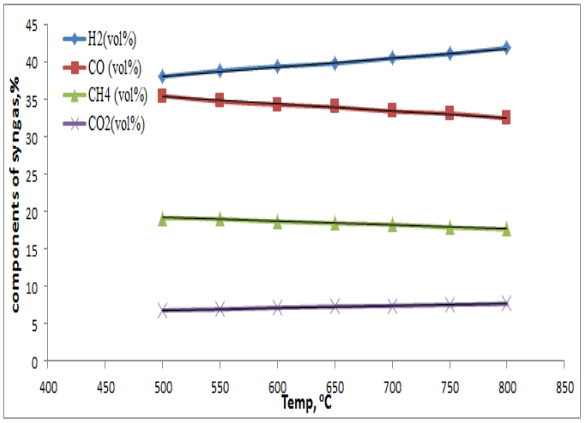
For 1:2 Dolomite-Sand mixture



For 1:1 Dolomite-Sand mixture

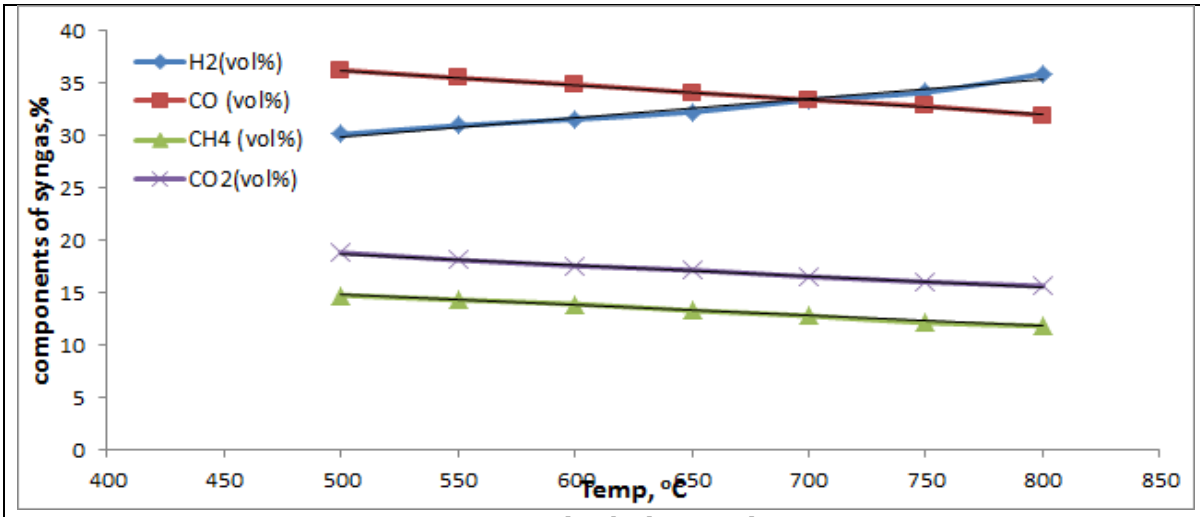


For 1:2 Red mud-Sand mixture

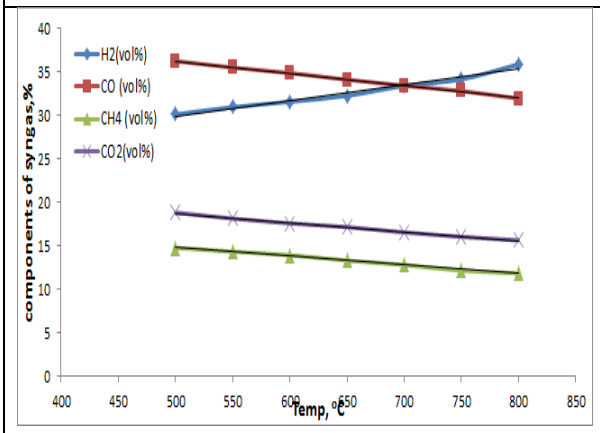


For 1:1 Red mud-Sand mixture

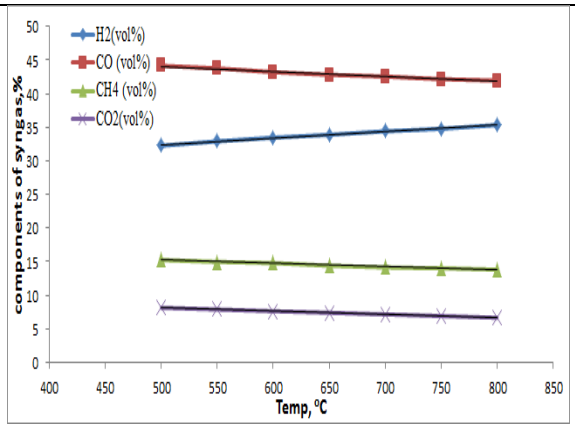
Fig.4.15: Comparison of effects of different bed materials for catalytic effects on yield of individual components of Syngas for sugarcane bagasse at S/B= 1.5, ER=0.25 and Feed Rate = 10kg/hr.



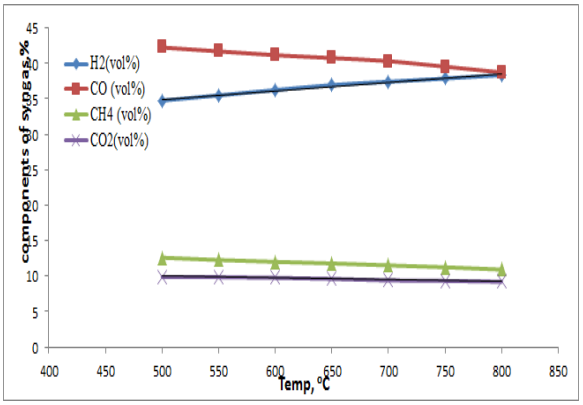
For Sand as bed material



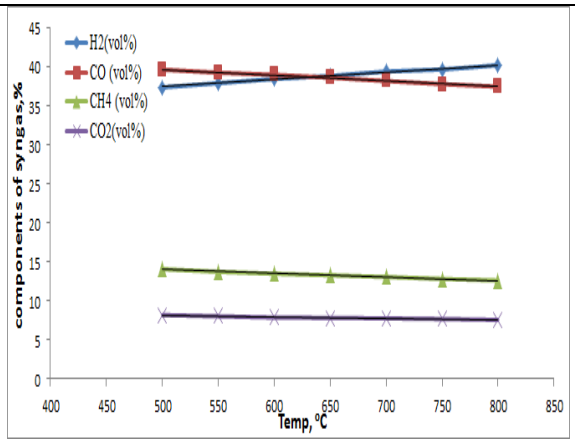
For 1:2 Dolomite-Sand mixture



For 1:1 Dolomite-Sand mixture

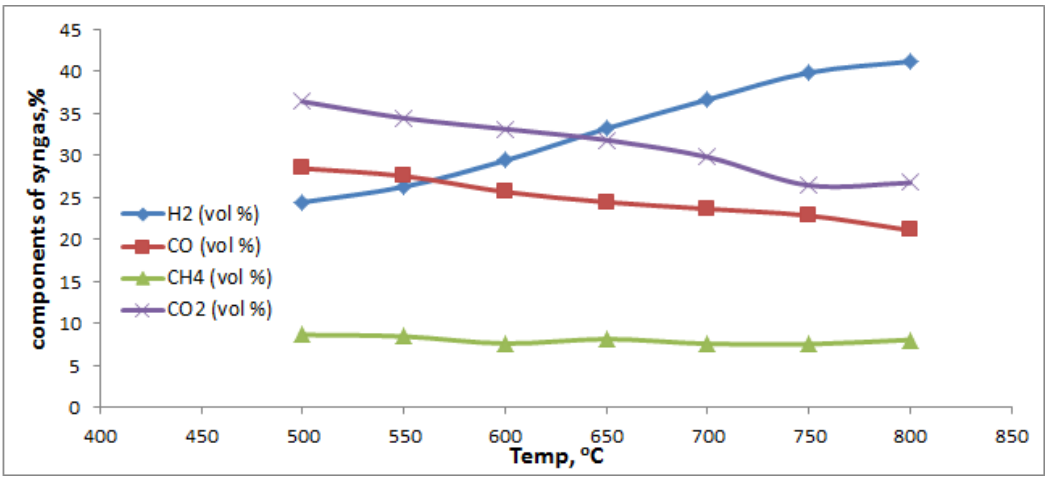


For 1:2 Redmud-Sand mixture

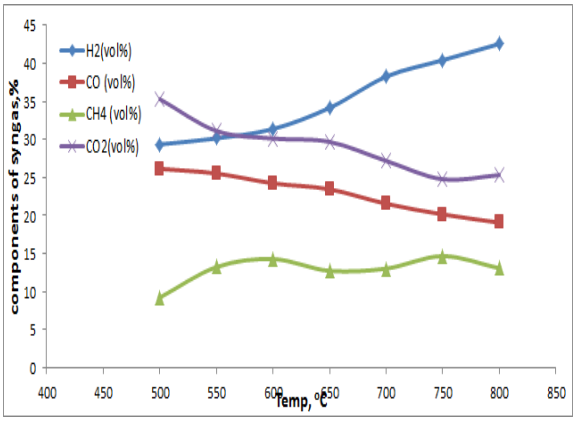


For 1:1 Redmud-Sand mixture

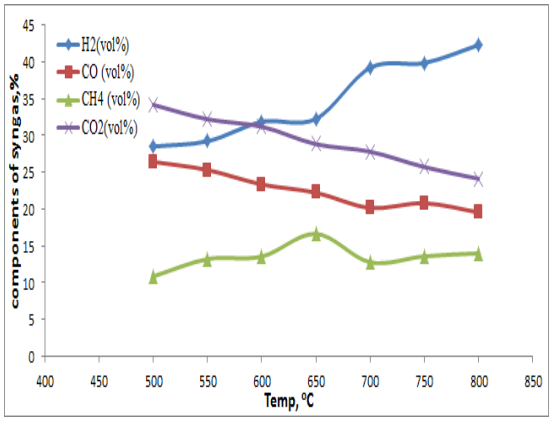
Fig.4.16: Comparison of effects of different bed materials for catalytic effects on yield of individual components of Syngas for coconut coir at S/B= 1.5, ER=0.25 and Feed Rate = 10kg/hr



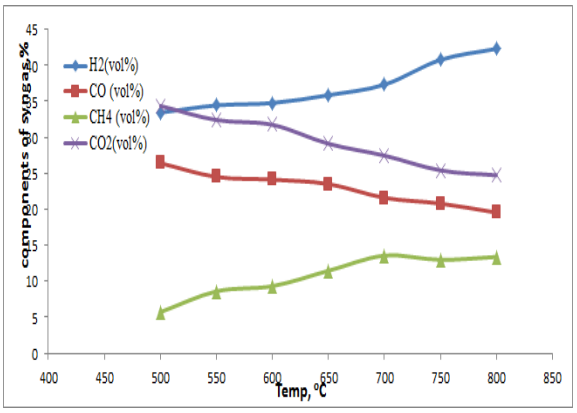
For Sand as bed material



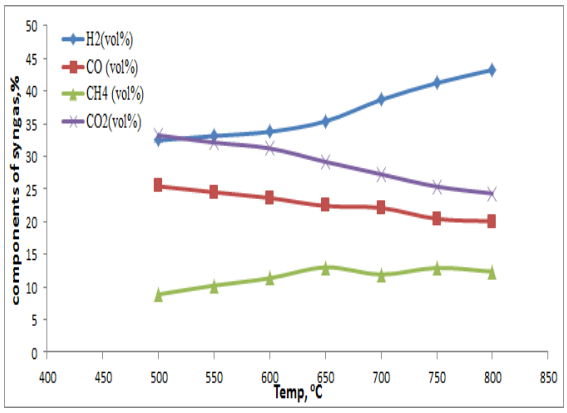
For 1:2 Dolomite-Sand mixture



For 1:1 Dolomite-Sand mixture

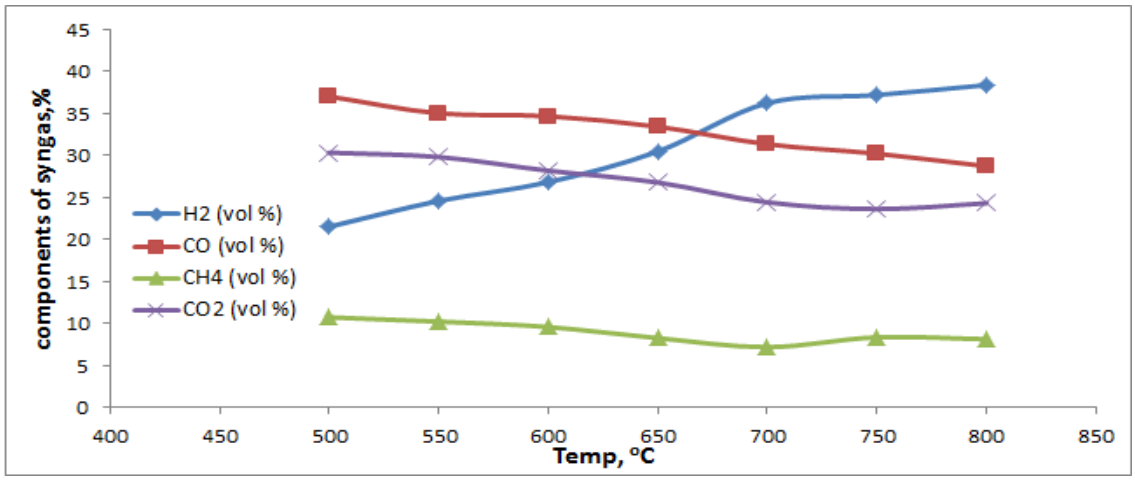


For 1:2 Redmud-Sand mixture

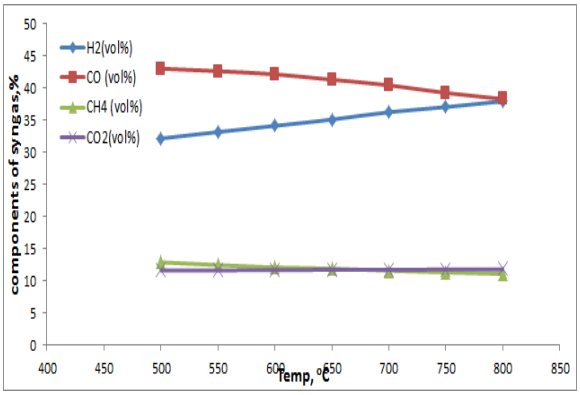


For 1:1 Redmud-Sand mixture

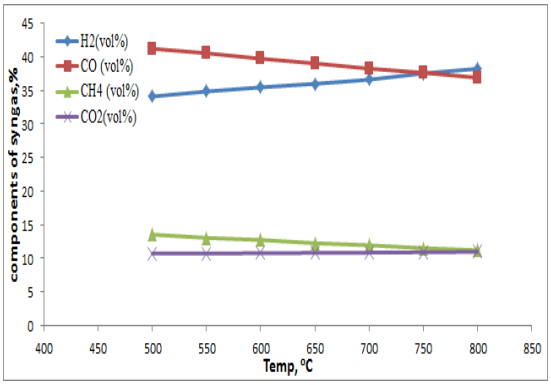
Fig.4.17: Comparison of effects of different bed materials for catalytic effects on yield of individual components of Syngas for wood chips at S/B= 1.5, ER=0.25 and Feed Rate = 10kg/hr.



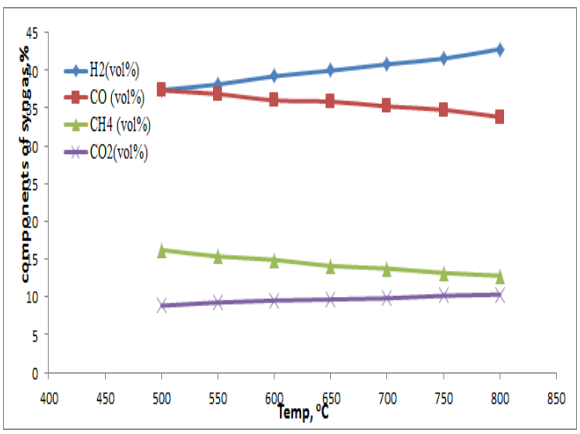
For Sand as bed material



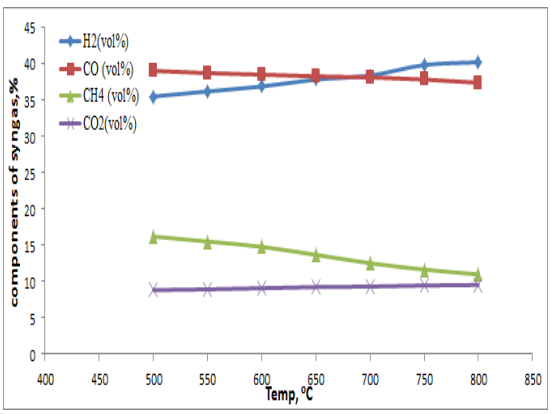
For 1:2 Dolomite-Sand mixture



For 1:1 Dolomite-Sand mixture

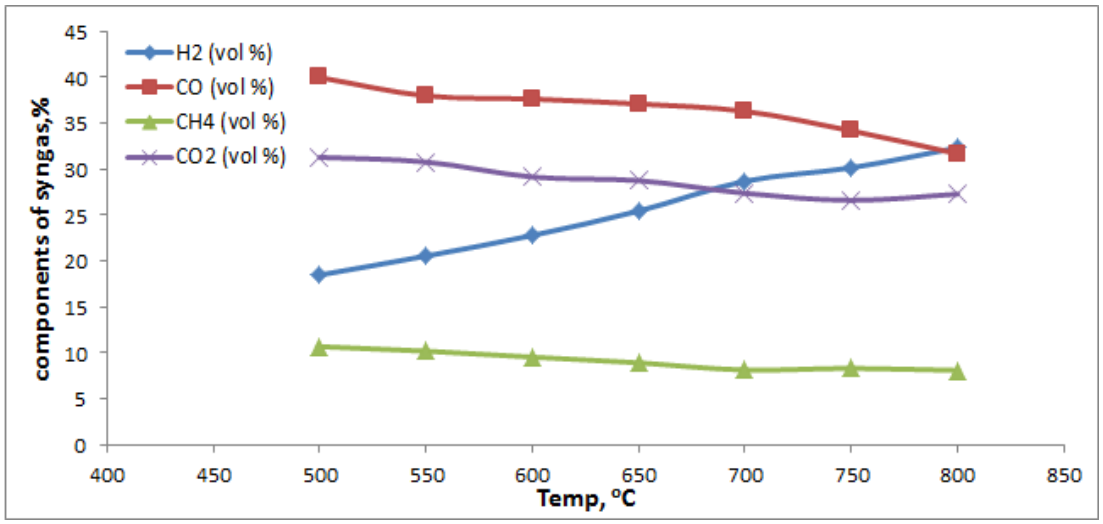


For 1:2 Redmud-Sand mixture

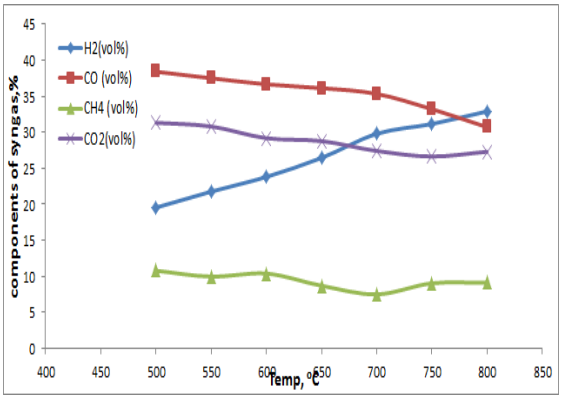


For 1:1 Redmud-Sand mixture

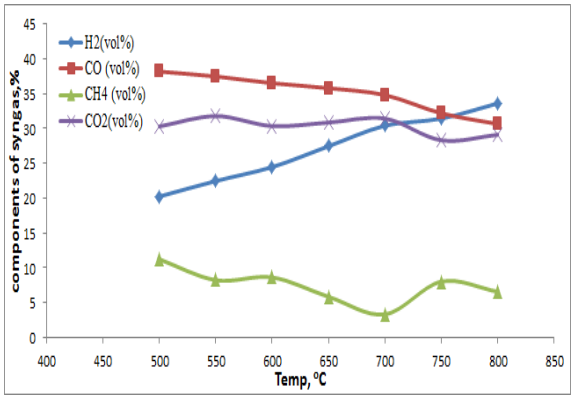
Fig.4.18: Comparison of effects of different bed materials for catalytic effects on yield of individual components of Syngas for rice husk at S/B= 1.5, ER=0.25 and Feed Rate = 10kg/hr.



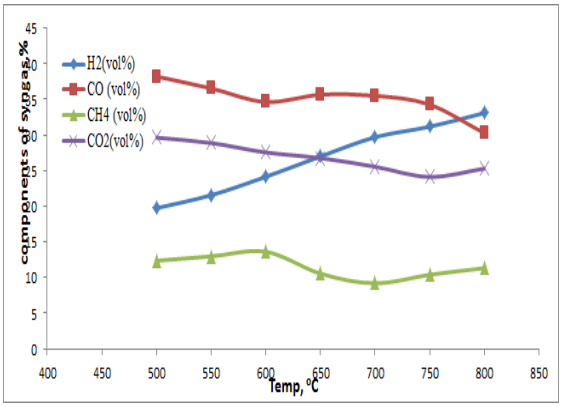
For Sand as bed material



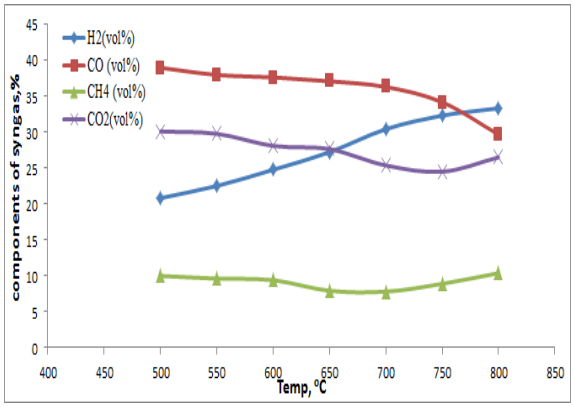
For 1:2 Dolomite-Sand mixture



For 1:1 Dolomite-Sand mixture

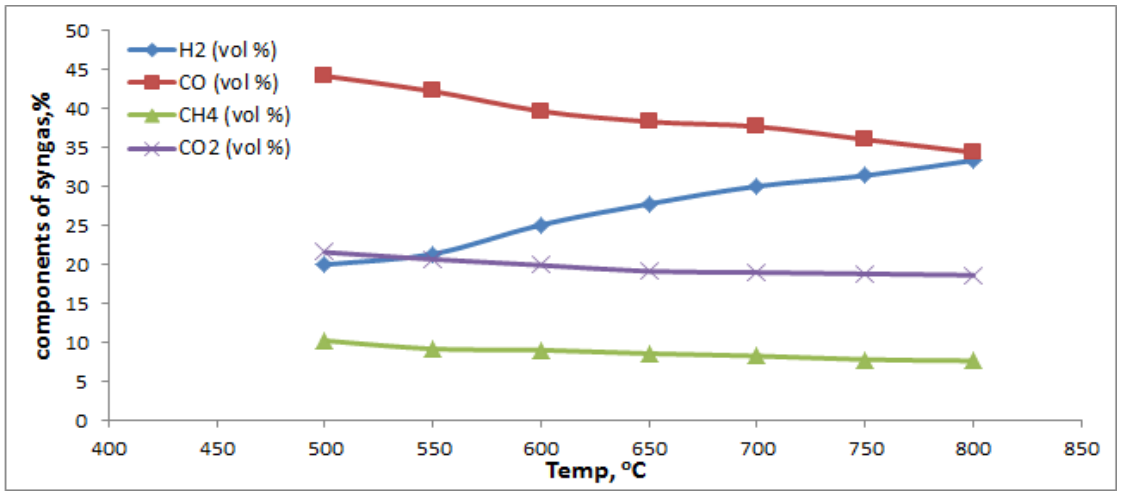


For 1:2 Redmud-Sand mixture

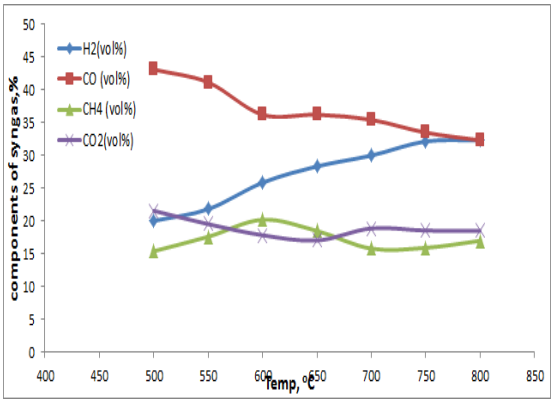


For 1:1 Redmud-Sand mixture

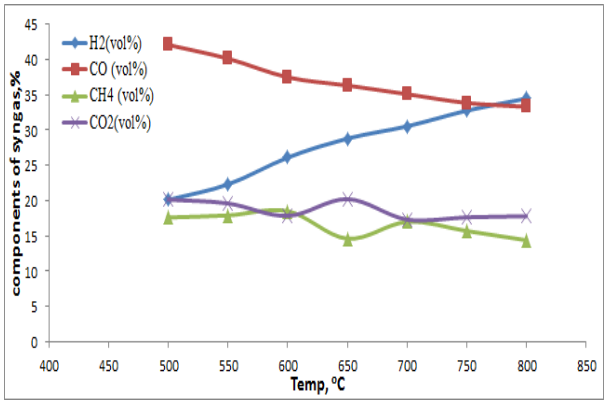
Fig.4.19: Comparison of effects of different bed materials for catalytic effects on yield of individual components of Syngas for rice straw at S/B= 1.5, ER=0.25 and Feed Rate = 10kg/hr.



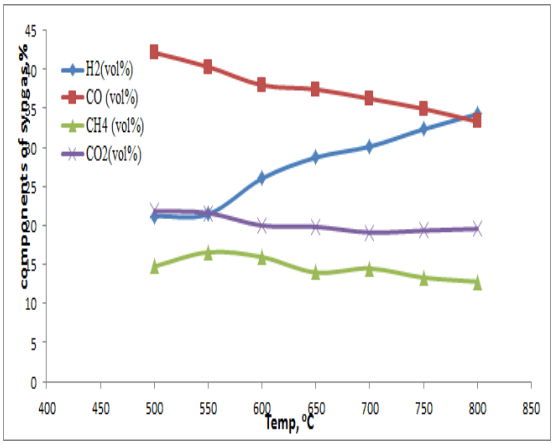
For Sand as bed material



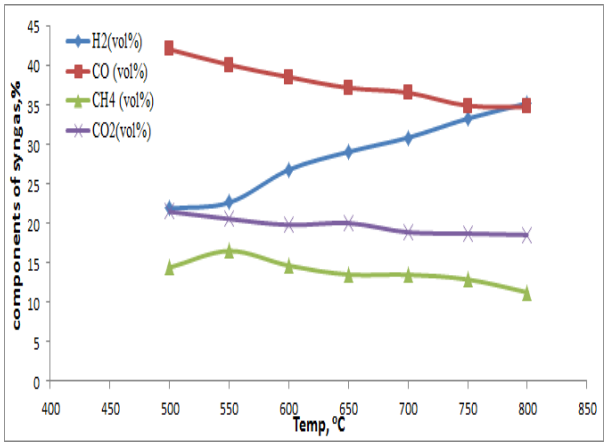
For 1:2 Dolomite-Sand mixture



For 1:1 Dolomite-Sand mixture



For 1:2 Redmud-Sand mixture



For 1:1 Redmud-Sand mixture

Fig.4.20: Comparison of effects of different bed materials for catalytic effects on yield of individual components of Syngas for saw dust at S/B= 1.5, ER=0.25 and Feed Rate = 10kg/hr.

Chapter-V

CFD Simulation

CFD simulation has been carried out with ANSY'S FLUENT -15 for bed hydrodynamics and reaction kinetics along with the temperature distribution within the Fluidized Bed Gasifier. Six different biomass samples are considered for the present work. Only one biomass sample has been shown in the main body of thesis and five other samples are reported in Appendix-III. In the present work, an Eulerian granular multiphase model is adopted where gas and solid phases are all treated as continua, interpenetrating and interacting with each other in the computational domain [74]. Fluid-fluid and fluid-solid (granular) multiphase flows are not distinguished. Any number of secondary phases can be modeled provided sufficient memory is available. A granular phase is simple and involves at least one phase as a granular phase. The pressure field is assumed to be shared by all the three phases in proportion to their volume fractions. Kinetic theory of granular flows is applied for determining solid-phase shear and bulk viscosities. It is planned to establish a simulation model to study thermal flow and gasification process using a fluidized bed gasifier by Eulerain multi-phase approach for which thermo-flow behavior with solids within the gasifier with no reactions has been considered. Therefore thermal flow for the gasification process is only analysed with 'Thermo-flow behavior with no reactions'. This thermo flow behavior is discussed in the next chapter with contour plots for different biomass samples

5.1 Governing Equations

5.1.1 Volume Fraction Equation

Volume fractions represent the space occupied by each phase, and the laws of conservation of mass and momentum are satisfied by each phase individually. The derivation of the conservation equations can be obtained by averaging the local instantaneous balance for each of the phases or by using the mixture theory approach.

The volume of phase q, V_q is defined by

$$V_q = \int_0^v \alpha_q \, . dV \tag{5.1}$$

Where

$$\sum_{q=1}^{n} \alpha_q = 1 \tag{5.2}$$

The effective density of phase q is calculated as

$$\rho_q = \alpha_q . \rho_q \tag{5.3}$$

Where ρ_q is the physical density of the phase q**.**

5.1.2 Conservation Equations

The motion of each phase is governed by respective mass, momentum and energy conservation equations.

(a) Conservation of mass:

The Continuity equation for phase q is

$$\frac{\partial}{\partial t}(\alpha_q \rho_q) + \nabla.(\alpha_q \rho_q \vec{v}_q) = \sum_{p=1}^{n}(\dot{m}_{pq} - \dot{m}_{qp}) + S_q \tag{5.4}$$

Where

$\vec{v}_q$ = Velocity of phase q

$\dot{m}_{pq}$ =the mass transfer from phase q to phase p

$\dot{m}_{qp}$ = the mass transfer from phase pto phase q

S_q = the source term of phase q

The right-hand side of Equation (5.4) is zero. This is because the net mass transfer from one phase to another is zero and the source term is considered by default zero except for the constant user-defined boundary conditions. Thus we have the following continuity equations:

(i) **For gas phase:**

$$\frac{\partial}{\partial t}(\alpha_g \rho_g) + \nabla.(\alpha_g \rho_g \vec{v}_g) = 0 \tag{5.5}$$

(ii) **For solid phase:**

$$\frac{\partial}{\partial t}(\alpha_s \rho_s) + \nabla.(\alpha_s \rho_s \vec{v}_s) = 0 \tag{5.6}$$

(b) Conservation of momentum:

Newton's second law of motion states that the change in momentum equals the sum of forces on the domain.

The momentum equation for phase q yields

$$\frac{\partial}{\partial t}\left(\alpha_q \rho_q \vec{v}_q\right) + \nabla.\left(\alpha_q \rho_q \vec{v}_q \vec{v}_q\right) = -\alpha_q \nabla.p + \nabla.\bar{\bar{\tau}}_q + \alpha_q \rho_q \vec{g} + \sum_{p=1}^{n}\left(\vec{R}_{pq} + \dot{m}_{pq}\vec{v}_{pq} - \dot{m}_{qp}\vec{v}_{qp}\right) + \left(\vec{F}_q + \vec{F}_{lift,q} + \vec{F}_{vm,q}\right) \quad (5.7)$$

Where, $\bar{\bar{\tau}}_q$ is the q[th]phase stress-strain tensor

$$\bar{\bar{\tau}}_q = \alpha_q \mu_q\left(\nabla.\vec{v}_q + \nabla.\vec{v}_q^{T}\right) + \alpha_q\left(\lambda_q - \frac{2}{3}\mu_q\right)\nabla.\vec{v}_q\bar{\bar{I}} \quad (5.8)$$

Where, μ_q = the shear viscosity of phase q, λ_q=the bulk viscosity of phase q,

$\vec{F}_q$ = an external body force of phase q, $\vec{F}_{lift,q}$ = a lift force of phase q,

$\vec{F}_{vm,q}$= a virtual mass force of phase q, $\vec{R}_{pq}$ = an interaction force between phase p and q

p= pressure shared by all phases, $\vec{v}_{pq}$ is the interphase velocity and is defined as follows.

If $\dot{m}_{pq}$ >0(i.e., phase p mass is being transferred to phase q),$\vec{v}_{pq} = \vec{v}_p$; If $\dot{m}_{pq} < 0$ (i.e., phase q mass is being transferred to phase p),$\vec{v}_{pq} = \vec{v}_q$. Similarly if $\dot{m}_{qp} > 0$, then$\vec{v}_{qp} = \vec{v}_q$; if $\dot{m}_{qp} < 0$, then$\vec{v}_{qp} = \vec{v}_p$.

The $\vec{F}_{vm,q}$,virtual mass force and the lift force $\vec{F}_{lift,q}$are considered zero by default. The equation 5.7 must be closed with appropriate expressions for the interphase force. The program uses a simple interaction term, in the following form:

$$\sum_{p=1}^{n}\vec{R}_{pq} = \sum_{p=1}^{n} Kpq\left(\vec{v}_p - \vec{v}_q\right)$$

Where Kpq (= Kqp) is the interphase momentum exchange coefficient.

Thus considering the above and$\dot{m}_{pq} = \dot{m}_{qp} = 0$, the general equations take the following form for the gas and solid phases.

(i) For gas phase:

$$\frac{\partial}{\partial t}\left(\alpha_g \rho_g \vec{v}_g\right) + \nabla.\left(\alpha_g \rho_g \vec{v}_g \vec{v}_g\right) = -\alpha_g \nabla.p + \nabla.\bar{\bar{\tau}}_g + \alpha_g \rho_g \vec{g} + Ksl\,(\vec{v}_g - \vec{v}_s) \quad (5.9)$$

(ii)For Solid phase**:**

$$\frac{\partial}{\partial t}\left(\alpha_s \rho_s \vec{v}_s\right) + \nabla.\left(\alpha_s \rho_s \vec{v}_s \vec{v}_s\right) = -\alpha_s \nabla.p + \nabla.\bar{\bar{\tau}}_s + \alpha_s \rho_s \vec{g} + Ksl\,(\vec{v}_g - \vec{v}_s) \quad (5.10)$$

(c)Conservation of Energy:

To describe the conservation of energy in Eulerian multiphase applications, a separate enthalpy equation is written for each phase:

$$\frac{\partial}{\partial t}\left(\alpha_q \rho_q h_q\right) + \nabla \cdot \left(\alpha_q \rho_q \overrightarrow{u_q} h_q\right) = \alpha_q \frac{\partial p_q}{\partial t} + \bar{\bar{\tau}} : \nabla \overrightarrow{u_q} - \nabla \overrightarrow{q_q} + S_q + \sum_{p=1}^{n} \left(Q_{pq} + \dot{m}_{pq} h_{pq} - \dot{m}_{qp} h_{qp} \right) \quad (5.11)$$

Where, h_q = the specific enthalpy of the phase "q", $\overrightarrow{q_q}$= the heat flux of the phase "q", Sq= a source term that includes sources of enthalpy, Q_{pq} = the intensity of heat exchange between the phase "p" and "q", h_{pq} = the inter-phase enthalpy.

5.2 Interphase Exchange Coefficient

5.2 .1 Fluid-solid Exchange Coefficient

The fluid-solid exchange co-efficient K_{sl} can be written in the following general form:

$$K_{sl} = \frac{\alpha_s \rho_s f}{\tau_s} \quad (5.12)$$

Where f is defined differently for the different exchange co-efficient model and τ_s, the particulate relaxation time is expressed as follows:

$$\tau_s = \frac{\rho_s d_s^2}{18\,\mu_l} \quad (5.13)$$

where d_s are the diameter of the particles of phase *s*. All definition of f includes a drag function *(*C_D*)* that is based on the relative Reynolds number *(*Re_s*)*.It is this drag function that differs among the exchange co-efficient models.

In the present study, Gidaspow model has been used, which is the combination of Wen and Yu model and the Ergun equation When$\alpha_l > 0.8$, the fluid solid exchange coefficient K_{sl} is of the following form:

$$K_{sl} = \frac{3}{4} C_D \frac{\alpha_s \alpha_l \rho_l |\vec{v}_s - \vec{v}_l|}{d_s} \alpha_l^{-2.65} \quad (5.14)$$

Where, $C_D = \frac{24}{\alpha_l Re_s}[\,1 + 0.15(\alpha_l Re_s)^{0.687}\,]$ (5.15)

Where Re_s is defined as

$$Re_s = \frac{\rho_l d_s |\vec{v}_s - \vec{v}_l|}{\mu_l} \quad (5.16)$$

l is the l^{th} fluid phase, *s*is for the s^{th} solid phase particles and d_s is the diameter of the s^{th}solid phase particles

When$\alpha_l \leq 0.8$,$K_{ls}$is written as

$$K_{ls} = \frac{3(1+e_{ls})\left(\frac{\pi}{2}+C_{fr,ls}\frac{\pi^2}{8}\right).\alpha_s\rho_s\alpha_l\rho_l(d_l+d_s)^2 g_{0,ls}|\vec{v}_l-\vec{v}_s|}{2\pi(\rho_l d_l^3+\rho_s d_s^3)} \quad (5.17)$$

Where, e_{ls} = the specific enthalpy of the phase "q", $C_{fr,ls}$ = the coefficient of friction between the l^{th} and s^{th}solid phase particles, d_l = diameter of the particle of solid l, $g_{0,ls}$ = the radial distribution coefficient.

5.2.2 Solid-solid Exchange Coefficient

The symmetric Syamlal model [78] is recommended for a pair of solids where the solid-solid exchange coefficient K_{ls} has the following form:

$$K_{ls} = \frac{3(1+e_{ls})+\left(\frac{\pi}{2}+C_{fr,ls}\frac{\pi^2}{8}\right)\alpha_s\rho_s\alpha_l\rho_l(d_l+d_s)^2 g_0 l_s}{2\pi(\rho_l d^3{}_l+\rho_s d^3{}_s)}|\vec{v}_s-\vec{v}_l| \quad (5.18)$$

Where, e_{ls} = the restitution coefficient
$C_{fr,ls}$ = the coefficient of friction between the $l^{th}$and s^{th} solid-phase particles ($C_{fr,ls}$ = 0),
d_l = the diameter of the particles of solid l, $g_0 l_s$ =the radial distribution coefficient.

5.3 Solid Pressure

For granular flow in the compressible regime (i.e. where the solid volume fraction is less than its maximum allow value), a solid pressure is calculated independently and used for the pressure gradient term ($\nabla . p_s$) in the granular-phase momentum equation. Because a Maxwellian velocity distribution is used for the particles, a granular temperature is introduced into the model which appears in the expression for the solid pressure and viscosities. The solid pressure is composed of a kinetic term and a secondary term due to particle collisions.

$$p_s = \alpha_s\rho_s\Theta_s + 2\rho_s(1+e_{ss})\alpha_s^2 g_{0,ss}\Theta_s \quad (5.19)$$

Where, e_{ss} = the co-efficient of restitution for particle collisions,
$g_{0,ss}$ = the radial distribution function, Θ_s = the granular temperature

The granular temperature Θ_s is proportional to the kinetic energy of the fluctuating particle motion. In ANSYS FLUENT a default value of 0.9 for Θ_s is used and can be adjusted to suit the particle type. The function $g_{0,ss}$ is a distribution function that governs the transition from the "compressible" condition with $\alpha_s < \alpha_{s,max}$ (where the spacing between the solid particles can continue to decrease) to incompressible condition with $\alpha = \alpha_{s,max}$ (where there is no further decrease in space). The default value for$\alpha_{s,max}$ is taken as0.63.

5.4 Radial Distribution Function

The radial distribution function g_o is a correction factor that modifies the probability of collision between grains when the solid granular phase becomes dense. This function may also be interpreted as the non-dimensional distance between spheres as mentioned below:

$$g_o = \frac{s + d_p}{s} \tag{5.20}$$

where s= the distance between grains and d_p = the diameter of particle.

For a dilute solid phase, $s \to \infty$ and therefore $g_o \to 1$. In the limit when solid phase contact, $s \to 0$ and $g_o \to \infty$. For a solid phase,

$$g_o = [1 - (\frac{\alpha_s}{\alpha_{s,max}})^{1/3}]^{-1} \tag{5.21}$$

5.5 Solid Shear Stresses

The solid shear stresses contain shear and bulk viscosities arising from particle momentum exchange due to translation and collision. A frictional component of viscosity can also be included to account for the viscous-plastic transition that occurs when particle of solid phase reach the maximum solid volume fraction. The collision and kinetics parts and the optional frictional part are added to give the solid shear viscosity as expressed below.

$$\mu_s = \mu_{s,col} + \mu_{s,kin} + \mu_{s,fr} \tag{5.22}$$

5.5.1 Collision Viscosity:

The collisional part of the shear viscosity is modeled as mentioned below[84,85]:

$$\mu_{s,col} = \frac{4}{5}\alpha_s \rho_s d_s g_{o,ss}(1 + e_{ss})(\frac{\Theta_s}{\pi})^{1/2}\alpha_s \tag{5.23}$$

5.5.2 Kinetic Viscosity:

The kinetic part of the shear viscosity is modeled in the following from[85]

$$\mu_{s,kin} = \frac{\alpha_s d_s \rho_s \sqrt{\Theta_s \pi}}{6(3 - e_{ss})}\left[1 + \frac{2}{5}(1 + e_{ss})(3e_{ss} - 1)\alpha_s g_{o,ss}\right] \tag{5.24}$$

5.5.3 Bulk Viscosity:

The bulk viscosity accounts for the resistance of the granular particle to compression and expansion. It has the following form [86].

$$\lambda_s = \frac{4}{3}\alpha_s \rho_s d_s g_{o,ss}(1+e_{ss})\left(\Theta_s/\pi\right)^{\frac{1}{2}} \quad (5.25)$$

5.5.4 Frictional Viscosity:

In dense flow at low shear, where the secondary volume fraction for a solid phase approaches the packing limit, the generation of stress is mainly due to friction between particles. In the present work, Schaeffer's expression for frictional viscosity as mentioned below is considered.

$$\mu_{s.fr} = \frac{p_s \sin\phi}{2\sqrt{I_{2D}}} \quad (5.26)$$

Where, $p_s$is the solids pressure,ϕ is the angle of internal friction, and I_{2D} is the second invariant of the deviatoric stress tensor.

5.5.5 Granular Temperature

The granular temperature for the s[th]solids phase is proportional to the kinetic energy of random motion of particles. The transport equation derived from kinetic theory takes the following form.

$$\frac{3}{2}\left[\frac{\partial}{\partial t}(\rho_s \alpha_s \Theta_s) + \nabla.(\rho_s \alpha_s \vec{v}_s \Theta_s)\right] = \left(-p_s\bar{\bar{I}} + \bar{\bar{\tau}}_s\right):\nabla.\vec{v}_s + \nabla.\left(K_{\Theta_s}\nabla.\Theta_s\right) - \Upsilon_{\Theta_s} + \quad \Phi_{ls} \quad (5.27)$$

Where, $\left(-p_s\bar{\bar{I}} + \bar{\bar{\tau}}_s\right):\nabla.\vec{v}_s$ =the generation of energy by solid stress tensor

$K_{\Theta_s}\nabla.\Theta_s$= the diffusion of energy ($K_{\Theta_s}$is the diffusionco-efficient)

Υ_{Θ_s}= the collisional dissipation of energy

Φ_{ls} = the energy exchange between the *l[th]*phase or solid phase and thes[th]solid phase

$K_{\Theta_s}.\nabla.\Theta_s$describe the diffusive flux of granular energy. The diffusion co-efficient for granular energy, $K_{\Theta_s}$is given by

$$K_{\Theta_s} = \frac{15\, d_s \rho_s \alpha_s \sqrt{\Theta_s \pi}}{4(41-33\eta)}\left[1 + \frac{12}{5}\eta^2(4\eta-3)\alpha_s g_{0,ss} + \frac{16}{15\pi}(41-33\eta)\eta\alpha_s g_{0,ss}\right. \quad (5.28)$$

Where,$\eta = \frac{1}{2}(1+e_{ss})$

The collisional dissipation of energy, Y_{Θ_s}, represents the rate of energy dissipation within the s^{th}solid phase due to collision between particles. This term is represented by the following expression:

$$\Upsilon_{\Theta_m} = \frac{12(1-e_{ss}^2)g_{0,ss}}{d_s\sqrt{\pi}} . \rho_s \alpha_s^2 \Theta_s^{3/2} \quad (5.29)$$

The transfer of the kinetic energy of random fluctuations in particle velocity from the s^{th}solid phase to the l^{th} fluid or solid phase is represented by ϕ_{ls} which is written as

$$\phi_{ls} = -3K_{ls}\Theta_s \quad (5.30)$$

5.6 Turbulence Model

To describe the effect of turbulent fluctuations of velocities in a multiphase flow, large numbers of terms are to be modeled in the momentum equations and this make the modeling of turbulence in multiphase simulations extremely complex. There are three methods for modeling turbulence in multiphase flow. These are mixture turbulence model, dispersed turbulence model and turbulence model for each phase. In the present work dispersed turbulence model is applied.

5.6.1 K − ε Dispersed Model:

This model is applicable only when there is clearly one primary continuous phase and rest are dispersed dilute secondary phases. In this case, interparticle collisions are negligible and the dominant process in the random motion of the secondary phase is the influence of the primary phase turbulence. Fluctuating quantities of the secondary phases can therefore be defined in term of the mean characteristics of the primary phase and the ratio of the mean particle relaxation time and eddy-particle relaxation time.

Turbulence in the Continuous Phase

The eddy viscosity model is used to calculate average fluctuation quantities. The Reynolds stress tensor for continuous phase, q takes the following form:

$$\bar{\bar{\tau}}_q'' = -\frac{2}{3}\left(\rho_p k_q + \rho_q v_{t,q} . \nabla . \vec{U}_q\right)\bar{\bar{I}} + \rho_q v_{t,q}\left(\nabla . \bar{\bar{U}}_q + \nabla . \vec{U}_q^T\right) \quad (5.31)$$

Where, $\vec{U}_q$ is the phase-weighted velocity

The turbulent viscosity $\mu_{t,q}$ is written in term of the turbulent kinetic energy of phase q as follows: expression:

$$\mu_{t,q} = \rho_q C_\mu \frac{k_q^2}{\varepsilon_q} \quad (5.32)$$

The characteristic time of the energetic turbulence eddies is defined as:

$$\tau_{t,p} = \frac{3}{2} C_\mu \frac{k_q}{\varepsilon_q} \tag{5.33}$$

Where, ε_q is the dissipation rate and $C_\mu = 0.9$.

The length scale of the turbulent eddies is written as:

$$L_{t,q} = \sqrt{\frac{3}{2}} C_\mu \frac{k_q^{3/2}}{\varepsilon_q} \tag{5.34}$$

Turbulent predictions are obtained from the modified $K-\varepsilon$ model as follows:

$$\frac{\partial}{\partial t}(\alpha_q \rho_q k_q) + \nabla.(\alpha_q \rho_q \vec{U}_q k_q) = \nabla.\left(\alpha_q \frac{\mu_{t,q}}{\sigma_k} \nabla k_q\right) + \alpha_q \rho_q \varepsilon_q + \alpha_q \rho_q \Pi_{k_q} \tag{5.35}$$

And $\frac{\partial}{\partial t}(\alpha_q \rho_q \varepsilon_q) + \nabla.(\alpha_q \rho_q \vec{U}_q k_q)$

$$= \nabla.\left(\alpha_q \frac{\mu_{t,q}}{\sigma_\varepsilon} \nabla \varepsilon_q\right) + \alpha_q \frac{\varepsilon_q}{k_q}(C_{1\varepsilon} G_{k,q} - C_{2\varepsilon} \rho_q \varepsilon_q) + \alpha_q \rho_q \Pi_{\varepsilon_q} \tag{5.36}$$

Here Π_{k_q} and Π_{ε_q} represent the influence the dispersed phase on the continuous phase q, and $G_{k,q}$ is production of turbulence kinetic energy.

The term Π_{k_q} is derived from the instantaneous equation of the continuous phase and takes the following form:

$$\Pi_{kq} = \sum_{p=1}^{M} \frac{k_{pq}}{\alpha_q \rho_q}(k_{pq} - 2k_q + \vec{v}_{pq}.\vec{v}_{dr}) \tag{5.37}$$

Where, M represents the number of secondary phases.

Turbulence in the Dispersed Phase:

Time and length scale which characterize the motion are used to evaluate dispersion co-efficient correlation functions and the turbulent kinetic energy of each dispersed phase.

The characteristic relaxation time connected with inertial effects acting on a dispersed phase p is defined as:

$$\tau_{F,pq} = \alpha_p \rho_q K_{pq}^{-1}\left(\frac{\rho_p}{\rho_q} + C_v\right) \tag{5.38}$$

The Lagrangian integral time scale is calculated along the particle trajectories and is mainly affected by the crossing trajectories. This is defined as:

$$\tau_{t,pq} = \frac{\tau_{t.q}}{\sqrt{(1+C_\beta \xi^2}} \tag{5.39}$$

Where

$$\xi = \frac{|\vec{v}_{pq}|\tau_{t,q}}{L_{t,q}} \tag{5.40}$$

and

$$C_\beta = 1.8 - 1.35(\cos\theta)^2 \tag{5.41}$$

Where θ is the angle between the mean particle velocity and the mean relative velocity.
The ratio between these characteristic times is written as:

$$\eta_{pq} = \frac{\tau_{t,pq}}{\tau_{F,pq}} \tag{5.42}$$

Turbulence quantities for dispersed phase, par written as:

$$k_p = k_q \left(\frac{b^2+\eta_{pq}}{1+\eta_{pq}}\right) \tag{5.43}$$

$$k_{pq} = 2k_q \left(\frac{b+\eta_{pq}}{1+\eta_{pq}}\right) \tag{5.44}$$

$$D_{t,pq} = \frac{1}{3} k_{pq} \tau_{t,pq} \tag{5.45}$$

$$D_p = D_{t,pq} + \left(\frac{2}{3}k_p - b\frac{1}{3}k_{pq}\right)\tau_{F,pq} \tag{5.46}$$

$$b = (1 + C_v)\left(\frac{\rho_p}{\rho_q} + C_v\right) \tag{5.47}$$

C_v= 0.5 is the added mass coefficient.

5.7 Species Transport Equations

The mixing and transport of chemical species are modeled by solving the conservation equations describing convection, diffusion, and reaction sources for each of the component species. The species transport equations are solved by predicting the local mass fraction of each species, Yi, through the solution of a convection-diffusion equation for i^{th} species. The species transport equation in general form is given as:

$$\frac{\partial}{\partial t}(\rho Y_i) + \nabla \cdot (\rho \vec{v} Y_i) = -\nabla \cdot \vec{J}_i + R_i + S_i \tag{5.48}$$

R_i = the net rate of production of species i by chemical reaction
S_i=the rate of creation by addition from the dispersed phase
An equation of this form will be solved for N-1 species where N is the total number of fluid phase chemical species present in the system. Since the mass fraction of the species must sum to unity, the N^{th} mass fraction is determined as one minus the sum of the N-1 solved mass fractions, since the total mass fraction must sum to unity.

$\vec{J}_i$ is the diffusion flux of species i, which arises due to concentration gradients. Mass diffusion for laminar flows is given as:

$$\vec{J}_i = -\rho D_{i,m} \nabla Y_i \quad (5.49)$$

For turbulent flows, mass diffusion flux is given as

$$\vec{J}_i = -\left(\rho D_{i,m} + \frac{\mu_t}{Sc_t}\right) \nabla Y_i \quad (5.50)$$

Where Sc_t is the turbulent Schmidt number.

CFD modelling for the hydrodynamic studies, thermal aspects and reaction kinetics within fluidized bed gasifier are also carried out in this chapter. ANSYS FLUENT.15.0 is used for simulation where 2D segregated first order implicit unsteady solver is used for multiphase calculations. Standard k-ε dispersed Eulerian multiphase model with standard wall functions are used. Gas is taken as continuous phase while binary mixtures of solid particles are taken as dispersed phase. Interphase interaction formulations used are of Solid–Solid (Syamlal-Obrien-symmetric) and Solid-Gas (Gidaspow)types. Inert material sand has been used as the bed material in the present work. Biomass (sugarcane bagasse) has been considered as the feed sample in FBG. Air is used as the fluidizing medium which is supplied from bottom of the FBG.

(A) Studies on Bed Hydrodynamics*:*

Before studying the details of reactions in different zones of the gasifier, it is essential to know the bed behaviour first such as identification and characterization of the flow regimes and structures in FBG[74]. That is why CFD modelling for the hydrodynamic studies of fluidized bed gasifier has been carried out in this work. The studies on bed hydrodynamics in this report include solid volume fraction, pressure drop profile, velocity vector, particle size distribution. In the present work, a parameter study on a bubbling fluidized bed has been carried out where the effects of gas velocity on the flow dynamics are studied.

5.8 Model and Simulation Method

ANSYS FLUENT.15.0 is used for simulation where 2D segregated first order implicit unsteady solver is used for multiphase calculations. Standard k-ε dispersed Eulerian multiphase model with standard wall functions are used. Gas is taken as continuous phase while binary mixtures of solid particles are taken as dispersed phase. Interphase interaction formulations used are of Solid–Solid (Syamlal-Obrien-symmetric) and Solid-Gas (Gidaspow) types. Sand has been used as the bed material and sugarcane bagasse has been considered as

the feed sample in FBG. Air is used as the fluidizing medium which is supplied from bottom of the FBG.

5.8.1 Assumptions Made

For carrying out simulation on any process, certain assumptions are required for initializing the computational work. In the present work, isothermal non-reactive, unsteady state gas-solid system are considered as basic assumptions in cold model FB gasifier unit. Eulerian multi-fluid model is adopted where both gas and solid phases are treated as continua, inter-penetrating and interacting with each other everywhere in the computational domain. The single pressure field is assumed to be shared for all three phases, in proportion to their volume fractions. Gas phase has been modeled with k-ε turbulent model and solid phases have been modeled with the kinetic theory of granular flow. The motion of each phase is governed by their mass and momentum conservation equations.

5.8.2 Geometry and Mesh

The reactor used for the bubbling fluidized bed is based on the experimental set up used in laboratory. Fig. 5.1(a) shows geometry of the reactor with its dimensions. The bubbling bed zone has inner diameters of 0.15m and height of 1m. The free board area has inner diameters of 0.3m and height of 0.8m. The geometry is generated by using commercial software GAMBIT. After geometry creation, a uniform mesh has been generated. Structured meshing method is used for meshing the geometry. In this study, total of 16,346 cells and16782 numbers of nodes are employed for simulating Fluidized Bed Gasifier.

5.8.3 Phases and Materials

The case is simulated using three phases, which enter the gasifier through boundary conditions and interact by exchanging mass and momentum. These phases are discussed below.

1. Gas phase: This is the Primary phase which is used for simulating both the air inlet and the product gas outlet.
2. Sand: It is the secondary phase which represents the fluidizing bed material in this study. The sand is modelled granular and inert with a constant size of 385 μm and density 2650 kg/m^3, belonging to Geldart B group.
3. Sugarcane bagasse: It is also considered as secondary phase which simulates the fuel inlet of the gasifier. It is also considered to be a granular phase.

5.8.4 Boundary and Initial Conditions

The initial and boundary conditions for the gas phase and solid phase simulations as used for the geometry are shown in Fig. 5.2 (a). The simulation is assumed to be non-reactive and a cold flow fluidization system therefore the operating conditions for the present case are assumed to be 300K temperature and 1 atmospheric pressure.

Fig.-5.2 shows the geometry of FBG where both sand and rice husk are initially in static condition in the fluidized bed with 0.1m initial static bed height. Thus the velocity of solid particles is set at zero and the inlet gas velocity at the bottom of FB is assumed to be uniform along the axial direction. The pressure is not specified at the inlet because of the incompressible gas phase assumption (relatively low pressure drop system). At the outlet, only pressure boundary condition is specified. The boundary condition at the walls is assumed such that the tangential and normal velocities are zero. Such conditions are known as no-slip boundary conditions.

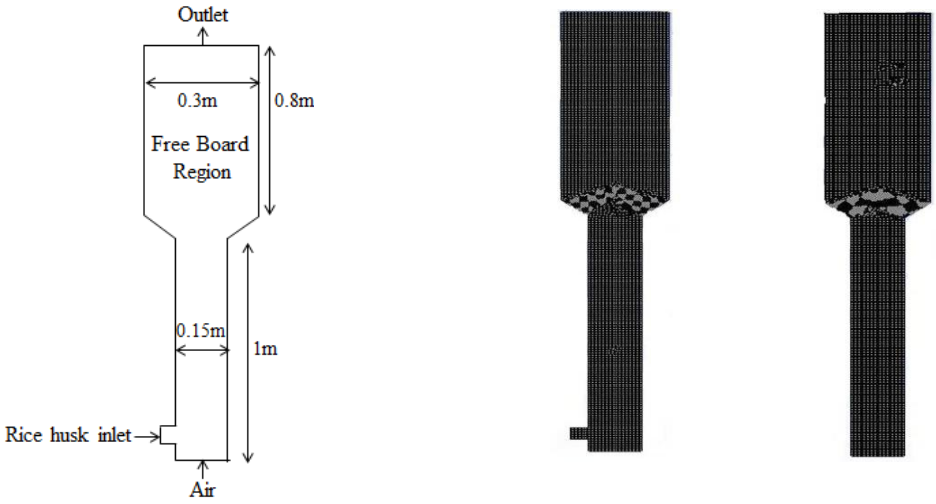


Fig.5. 1(a) Geometry of fluidized bed (b) 2-D Mesh1 (c) 2-D Mesh2

Table 5.1: Simulation model parameters used for gas and solid flow in a FBG

Property	Value	unit
Gas density(air)	1.2	kg/m^3
Gas viscosity	$1.7894*10^{-5}$	Pa.s
Biomass density	120	kg/m^3
Biomass particle diameter	10	mm
Density of inert solid(sand)	2650	kg/m^3
Diameter of sand	385	μm
Superficial gas velocity	0.2	m/s
Static bed height	0.1	m
Biomass inlet velocity	0.005	m/s

Restitution coefficient, e	0.9	-

5.8.5 Solution Techniques

The Phase Coupled SIMPLE method is chosen for pressure–velocity coupling. The second-order upwind scheme is used for discretization of momentum and turbulence kinetic energy.Turbulence dissipation rate and the first-order upwind scheme are used for discretization of volume-fraction equations. The time step of size =0.001s is taken for the solution to converge.Simulation model parameters used for gas and solid flow in a FBG are listed in Table-5.1.Under relaxation factors for different flow quantities are mentioned in Table- 5.2.

Table 5.2: Under relaxation factors for different flow quantities

Variable	Relaxation Factor
Pressure	0.5
Density	1
Body Force	1
Momentum	0.2
Volume Fraction	0.4
Granular Temperature	0.2
Energy	0.8

Volume fraction, velocity distribution and pressure distribution are analysed under hydrodynamic studies for gasification of different biomass samples in a fluidized bed gasifier. Contour plots of these profiles for different biomass samples are shown in Appendix-III. Contour plots for one sample are discussed in the next chapter under results and discussions.

Chapter-VI

Results and Discussions

Energy analysis for selected six different biomass samples being gasified in a FBG is carried out in this chapter. The gasifier performance with respect to these results of energy analysis is also discussed. Knowledge about various elements present in feed sample and product gas with their quantities is very much essential for calculation of chemical formula, carbon conversion efficiency and thermal efficiency for the gasifier handling any biomass sample. Biomass samples are characterized by the ultimate and proximate analysis. Product gas namely, syngas of different feed samples produced from the gasifier are analyzed with the help of the gas analyzer. Gasification operation is carried out in two steps. First step is for preheating of the biomass sample to the predefined temperature while in second step, biomass is gasified to produce the flue gas. During gasification, steam is pinched at a rate of 0.5kg per kg of fuel for which some energy is spent. Again feed sample needs to be sized and fluidized with the bed materials for which some energy requirement is also there. Thereforethe energy analysis is very much essential before making the process commercial. This chapter is described in two parts namely, results and discussions.

(A) Experimental Results and Discussions

Gasification is carried out by considering six different biomass samples viz. Rice husk, Rice straw, Wood chips, Saw dust, Coconut coir and Sugarcane bagasse. Sand, dolomite-sand mixture (1:1 & 1: 2 proportions) and red mud-sand mixture (1:1 & 1: 2 proportions) are used as bed materials. Same operating conditions are maintained during gasification for all the biomass samples. Temperature is varied within 500 to 1000^{o}C. Steam to biomass ratio is varied within 0.5 to 2.5 and equivalence ratio is varied within 0.15 to 0.35. Effect of these parameters on yield of hydrogen component of product gas is focused here in this chapter.

6.1 Chemical Formula of Biomass

The calculation of chemical formula is essential to determine the stoichiometric amount of air required for the combustion of the biomass sample [75]. Neglecting Nitrogen and Sulfur content, the formula of biomass should be C_xH_yOz. Considering Carbon as the base for the carbonaceous material, the chemical formula for the sample becomes CH_mO_n.

The chemical formula for rice husk, rice straw, saw dust, wood chips, sugarcane bagasse and coconut coir are calculated. The chemical formula for these biomass samples with and without N and S contents are shown below in Table - 6.1.

Table – 6.1: Chemical formula of biomass samples

Biomass Samples	Chemical formula of Biomass	
	With N, S	Without N, S
Rice husk	$CH_{1.55}O_{1.08}N_{0.02}S_{0.02}$	$CH_{1.55}O_{1.08}$
Rice Straw	$CH_{1.49}O_{1.19}N_{0.011}S_{0.0021}$	$CH_{1.49}O_{1.19}$
Saw Dust	$CH_{1.392}O_{0.8}N_{0.0037}S_{0.00057}$	$CH_{1.39}O_{0.8}$
Wood chips	$CH_{1.48}O_{0.74}N_{0.0042}S_{0.001}$	$CH_{1.48}O_{0.74}$
Sugarcane bagasse	$CH_{1.667}O_{0.787}N_{0.0038}S_{0.0042}$	$CH_{1.667}O_{0.787}$
Coconut coir	$CH_{1.589}O_{0.808}N_{0.0078}S_{0.0019}$	$CH_{1.589}O_{0.808}$

6.2 Energy Balance and Mass Balance Calculations

Based on the properties of the different biomass samples, the energy balance and mass balance calculations are carried out [1]. A sample calculation for sugarcane bagasse is given below while these calculations for other five samples are given in Appendix-II. With the help of material balance for hydrogen content, amount of hydrogen, the clean energy to be produced from each biomass sample has been estimated. With the help of material balance calculations for carbon content, the carbon conversion efficiency, cold gas efficiency and thermal efficiency for the gasifier are also calculated for different feed samples.

6.2.1 Sample calculation

The detailed calculation of energy requirement for gasification process is carried out as per the following.

With S/B ratio of 0.5, 0.5kg of steam is supplied for 1.0kg of fuel samples. So energy required for production of 0.5kg of steam will be

0.5*4.2*(100-30) +0.5*2260=1277kJ as per the formula $mc_p\Delta T + m\lambda$.

Thus for generation of 5.0 kg of steam (for 10.0kg of fuels) energy requirement will be 1277*10=12770 kJ

Energy requirement for preheating of bed material and fuel sample is 6969.275 kJ

Some energy is spent for sizing the biomass sample. Some energy is also spent for running the blower to supply air for fluidization of bed materials. As air supply is stopped after

around 10 minutes of fluidization, roughly 10000 kJ of energy can be considered for both the operations i.e. sizing of feed sample and supplying of air for fluidization.
Thus total energy input requirement for carrying out gasification of biomass sample can be calculated approximately as
6969.275 +12770 + 10000 = **29739.275 kJ** (= 8.26 kWh)
Time required for preheating is around 60minutes and operation time of gasification stage is around 30 minutes. During gasification it is not possible to distinguish between preheating and gasification periods. As 10 kg of feed sample is fed over a period of 1.0 hour, gasification might have started by the time last batch of feed sample reach the gasifier. Therefore total time of gasification operation may be taken as 60+30= 90minutes = 1.5 hours
Thus total power requirement can be calculated as total energy requirement /total time of operation. Thus, input power which is roughly same for all feed samples considered in the present study can be calculated as
Input power = 29739.275/1.5 = 19824.1833 kJ/h = 5.51 kJ/s = 5.51kW

Considering Sugarcane bagasse for the case study, energy output is going to be determined here in this section. Energy balance and mass balance calculations with the energy analysis report for five other samples are reported in **Appendix-II**. Energy analysis report for sugarcane bagasse has already been published in Energy Journal [87].

For Sugarcane Bagasse sample- A case studies:

Data:
Biomass Fuel - 10 kg/hr (92% dry)
Steam – 0.5 kg/kg fuel (corresponding to S/B ratio of 0.5)
Moisture is 8%
Composition of fuel on %dry basis
H_2 =6.2, C = 44.6, N_2 =0.2, O_2 = 46.84, S = 0.5
Product gas:
H_2 =12.38%, CH_4 = 3.05%, CO = 10.71%, CO_2 =8.25%, N_2 = 50.3% and O_2=15.3%
HHV=17.35 MJ/kg
From experiment air flow rate is noted to be = 0.00767m^3
Specific humidity of supply air be 0.01kg moisture per kg dry air

Mass Balance:

Basis: 1 kg of fuel with 0.08 kg moisture, 0.92kg dry fuel
Air: Flow rate = 0.00767m^3 (corresponding to Equivalence Ratio=0.2) for Feed rate of 10kg/hr

Thus air flow rate for 1kg/hr of feed is calculated as = 2.76kg/kg of fuel

Steam: 0.5 kg/kg of fuel (corresponding to S/B ratio of 0.5)

Air contains 76.9% N_2 by mass

So N_2 from air

= 0.769* 2.76 = 2.122 kg N_2/kg fuel

Total N_2 supplied by air and fuel which carry 0.2% N_2

= 2.12244+ 0.002 = 2.1244 kg N_2 /kg fuel

= 0.075943 kg mol/ kg fuel

We know product gas contains 50.3% N_2. Hence the amount of product gas produced

= 0.079543/ 0.503 = 0.151 kg mol gas / kg fuel =151 moles of gas/kg fuel.

Product Gas=0.151 kg mol / kg fuel = 13.294 m^3/kg fuel at 800^0C (3.3824m^3/kg fuel at STP)

Oxygen inflow to Gasifier:

Oxygen supplied to gasifier (with air) is

= 0.2315 * 2.76 = 0.639 kg/kg fuel

Steam supplied = 0.5 kg /kg fuel

So oxygen associated with steam supply

= 16/18 *0.5 = 0.44 kg /kg fuel

Moisture in fuel = 8%

Oxygen with moisture in fuel

= 0.08 * 16/18 = 0.071111kg/kg fuel

Oxygen with moisture in air supply

= 0.01 * 2.76 * 16/18 = 0.024533 kg/kg fuel

Oxygen in fuel =39.84% =0.3984kg/kg fuel

Total oxygen flow to gasifier with air, steam, moisture in fuel and air

= 0.6389 + 0.44444 + 0.0711 + 0.024333 + 0.3984

= 1.577184 kg/kg fuel=1.577 kg/kg fuel

H_2 Balance:

Total H_2 in flow to the gasifier with fuel, steam, moisture in fuel and moisture in air

= 0.08* 2/18 + 0.5 * 2/18 + 0.062*0.92 + 0.01* 2.76*2/18

= 0.12455111 kg/kg fuel

H_2 associated with H_2 & CH_4 in dry product gas

= (0.1238 + 0.0303 * 2) * 0.151

= 0.0278 kg mol/kg fuel

=0.0557 kg H_2 /kg fuel

Assuming all S converted to H_2S and removed by the gas cleaning system, hydrogen associated with H_2S in the raw product gas

$=0.005 * 2/32 = 3.125* 10^{-4}$ kg/kg fuel

Total hydrogen in the unclean dry product gas including that in H_2S

$=0.0556888 + 3.125 * 10^{-4}$

= 0.056kg H_2 /kg fuel

To find the moisture in the product gas we deduct the hydrogen in dry gas from the total hydrogen in flow obtained earlier using the hydrogen balance.

= Hydrogen in flow – Hydrogen in dry gas

=0.12455111- 0.056

= 0.0686 kg/kg fuel

Steam associated with this hydrogen in the gas

=0.0686 * 18/2 = 0.617 **kg /kg fuel**

Oxygen Balance:

Oxygen associated with CO, CO_2 in dry gas which have half a mol and 1 mol of oxygen respectively.

= (0.5 * 0.1071 + 0.0825+0.153) * 0.151

= 0.04364655 kg mol /kg fuel

= 0.04364655 * 32 = 1.397 kg /kg fuel

Oxygen associated with the steam in gas

= 0.61694829 * 16/18 =0.548 kg /kg fuel

Total Oxygen in gas = 1.397+0.548= 1.945**=1.9451 kg/kg fuel**

Here we note that this is slightly more than the oxygen in flow of **1.577 kg /kg fuel** calculated earlier. This must be due to measurement errors in the given data on fuel gas composition.

Carbon Balance:

Total carbon associated with CO, CO_2 and CH_4 in dry gas whose production rate has been computed earlier as 0.152 kg-mol / kg fuel is

= (0.1071 + 0.0825+0.0303) * 0.151

= 0.0332049kg mol/ kg

= 0.0332049 * 12 = 0.3984588**= 0.398 kg/kg fuel**

Carbon input is found from the composition of the fuel

= **0.446kg /kg fuel**

Carbon conversion efficiency

= (0.39846/0.446) * 100

= 0.8934=89.34%

Energy Balance:

Higher heating value (HHV) for different components of output gas

Components	Heats of Combustion, MJ/kg mol
CO	282.99
H_2	285.84
CH_4	890.36

Energy output with CO

=0.1071*0.151(kg mol CO/kg fuel)*282.99= 4.576543MJ/kg fuel

Energy output with CH_4

=0.0303*0.151(kg mol CO/kg fuel)*890.36= 4.073664MJ/kg fuel

Energy output with H_2

=0.1238*0.151(kg mol CO/kg fuel)*285.84= 5.343436MJ/kg

Total Energy output

=4.576543+ 4.073664+ 5.343436 **= 13.994 MJ/kg fuel**

Total energy input is 17. 35 MJ/kg fuel

Cold gas efficiency of the gasifier

=(13.994 / 17. 35)*100= 80.655 %

6.2.2 Proposed Energy Analysis Report

(a) Energy Input

Gasification process has two sub-processes. One is for preheating the biomass to the predefined temperature while other one is for the Gasification of biomass to produce the flue gas.

The energy required in the pre-heating stage will be approximately 2.5kWh and time required for preheating will be around 60minutes.

The energy required in the gasification stage will be approximately 1.5 kWh and operation time of gasification stage will be required around 30 minutes.

Amount of biomass burnt on a daily basis will be 10 kg/hour.

Hence total amount of energy utilized daily will be = [{(2.5+1.5)/ (90/60)}] =2.6667 kW (approx.)

During gasification steam is pinched at a rate of 0.5kg per kg of fuel. Thus some energy is spent for generating steam. Again feed sample needs to be sized. Thus the energy requirement is calculated as follows. With specific heats for sugarcane bagasse and sand as 0.836 and 0.795kJ/kg-^{0}C respectively, heat requirement to ignite the fuel sample and heat the bed material up to 600^{0}C. Is found to be = [0.836*10*+ 0.795*5]*(600-35) = 6969.275 kJ/h = 1.94 kW. Energy required for production of 0.5kg of steam will be 0.5*4.2*(100-30) +0.5*2260=1277kJ. Considering 10000 kJ of energy requirement for sizing of feed sample and supplying of air for fluidization, total energy input requirement for carrying out gasification of sugarcane bagasse can be calculated as = 6969.275 +1277*10 + 10000 =

29739.275 kJ (= 8.26 kWh). As total time of gasification operation is taken as 60+30= 90minutes and total power requirement will be calculated as = 29739.275/1.5 = 19824.18 kJ/h = 5.51 kJ/s = 5.51kW

(b) Energy output (heat energy extracted from the flue gases)

Energy output was calculated using the heating value of the flue gases.

As per material balance calculation it is found that 1kg of biomass sample produces 0.151kmol gas which occupies =0.151*22.4=3.3824m^3at STP. This implies that vol. of gas produced at 800^{0}C will be =3.382*(1073/273) =13.294 m^3, this is under ideal conditions. But the gas analyzer reads product gas as 10 m^3/kg of biomass samples. This difference in yield might be due to the fact that the gasifier is not 100% efficient for which all the biomass are not gasified completely and some losses of product gas might be there.

With the density of flue gas = 0.345kg/m^3 and mean specific heat at 800^{0}C as 3.735kJ/kg.^{o}K, heat of product gas can be calculated as

$\sum mc_p\Delta T$ **= (10*0.345)*3.735 * (800-35) =** 9857.6kJ or 2.74 kWh (approx.) since m= vol*density

Flow rate of the flue gases from the gasifier unit is observed to be 10 m^3/ hr. Composition of the flue gas for air gasification is as shown below in Table 6.2 (A).

Table 6.2 (A): Composition of flue gas obtained from sugarcane bagasse gasification

	O_2	CO	CO_2	CH_4	H_2	N_2
Mole or, Vol%	15.3	10.71	8.25	3.05	12.38	50.3
Flow rate(m^3/hr)	1.53	1.071	0.825	0.305	1.238	5.03

Temperature of the flue gases coming out is 800^{0}C. The heat energy processed by the flue gases was sapped out before calculating the net heating value of the flue gases. From the energy balance calculations it is found that 0.151kmol of producer gas is produced per kg of fuel sample. Using ideal gas law it is observed that

0.151kmol gas occupies =0.151*22.4=3.3824m^3at STP. This implies that vol. of gas at 800^{0}C will be =3.382*(1073/273) =13.294 m^3= 14m^3 (say).

As all the biomass is not gasified completely and some losses might be there, gasifier yield is considered to be 75%. Thus amount of gas yield from the gasifier comes 10 m^3/h (approx.).

The heat energy extracted from the flue gases:

(Density of flue gas =0.345kg/$m^3$and mean specific heat at 800^{0}C is 3.735KJ/kg.^{0}K)

Heat Energy: $\sum mc_p*(800-35) = (10*0.345)*3.735*765=12.886*765= 9857.599$ kJ

= 9.858 MJ/hr=2.738 kWhr/hr=**2.74 kW power(approx.)**

Energy expected to be obtained from the flue gases is mentioned below (Table-6.2(B)).

Table-6.2 (B): Heating value of major components of flue gas

Compound	Heating Value, (KJ/m^3)
CO	4074
CH_4	11605
H_2	3450

Output energy expected to be obtained from the different components of the flue gas is calculated as = (1.071 *4074) + (0.305*11605) + (1.238*3450) = 12173.88 kJ/hr = 12.17 MJ/hr

= 3.38166 kWhr/hr= 3.382 kW (**approx.)**

Total Energy output = 2.74+ 3.382 = 6.122kWh per kg of fuel

Net energy output= 6.122*10 – 8.26 = 52.96 kWh

10 kg of feed (dry basis) produces 52.96 kWh energy equivalent of synthesis gas (**52.96/1.5=35.3kW power**) and 10 $Nm^3$of gas per hour.

(c)Typical Operation Yields and Efficiencies

The operation results give the following yields and efficiencies.

1 kg of feed (dry basis) yielded 3.53 kW power or 5.296 kWh energy equivalent of synthesis gas.

1 kg of feed yielded 1.0 Nm^3 of gas with 12.3817% H_2, 3.05% CH_4, 10.71% CO, 8.25% CO_2, 15.3% O_2 and 50.3% N_2.

Hydrogen Yield= 0.056 kg/kg fuel = 56 gm of H_2 per kg of sugarcane bagasse

Power output is calculated by using the heating value of the flue gases. Flow rate of the flue gases from the gasifier unit is measured as 10 m^3/h. Composition of the flue gas produced from gasifier is determined by the gas analyzer (Table-6.2). Temperature of the flue gases coming out is found to be 800^0C. The heat energy processed by the flue gases is sapped out before calculating the net heating value of the flue gases.

Efficiency of the gasifier handling different biomass samples are calculated on the basis of energy balance. Total energy output is calculated as the sum of the energy output with CO, H_2 and CH_4 of each sample. Energy input is taken as the energy utilized for the gasification process i.e. 8.26kWh or 29739.275kJ. Thus it is calculated as follows.

Gasifier Efficiency for sugarcane bagasse = 6.122kWh/8.26kWh=0.74122 =74.22%

Thermal conversion efficiency of the gasifier for biomass can also be calculated using data given in Table 3.5 as per the following.

Thermal conversion efficiency = [(Net heating value indicated by gas analyzer)*(gas production rate)]/ [higher heating value of biomass sample]

Thus Thermal conversion efficiency for sugarcane bagasse = (2650kCal/m^3)*(1.4 m^3/kg of fuel)*0.0042/20 (since 1kCal=0.0042MJ).

Or, **Thermal conversion efficiency = 0.7791 = 77.91%**

Finally all the results related to gasification process with different biomass samples (i.e. cold gas efficiency, carbon conversion efficiency, amount of flue gas produced and proposed net energy output) are tabulated in Table-6.3. Carbon conversion efficiency, thermal conversion efficiency and efficiency of the gasifier calculated for different biomass samples are compared and shown in Table - 6.4.

Table 6.3: Final Result for Biomass gasification using a Fluidized Gasifier

Biomass sample	H_2-yield kg/kg of fuel	Carbon Conversion efficiency,%	Cold gas Efficiency, %	Fluegas,m^3/hr for 10kg/hr feed rate	Net Energyoutput ,kWhr/kg fuel
Rice husk	0.073874	93.13	82.08	11	6.21
Rice straw	0.06061	95.0	83.05	10	5.17
Saw dust	0.063914	77.76	88.32	11	5.92
Wood chips	0.058675	70.42	85.8	10	5.09
Sugarcane bagasse	0.056	89.34	80.655	10	5.296
Coconut coir	0.056682	82.3	75.686	10	5.30

Table-6.4: Comparison of efficiency of the gasifier with different types of biomass samples

Biomass sample	Carbon conv. efficiency, %	Thermal conv. efficiency, %	Gasifier efficiency, %	Deviation, %
Rice husk	93.13	79.71	85.18	-0.264
Rice straw	95.0	74.97	72.59	2.013
Saw dust	77.76	75.09	81.6	3.681
Wood chips	70.42	76.22	71.62	2.307
Sugarcane bagasse	89.34	77.91	74.22	-3.576
Coconut coir	82.3	74.26	74.16	0.536

6.3 Development of Correlations for Hydrogen Yield

Effect of system parameters on the yield of hydrogen is analyzed for different biomass samples. Thus attempt is made to develop correlations for yield of Hydrogen on the basis of dimensional analysis by relating experimental values of hydrogen yield obtained through gasification of biomass samples (as a component of product gas) to different system parameters viz. temperature (T), steam to biomass ratio (S/B), equivalence ration (ER) and density of bed materials (ρ_M) as follows (Eq. nos.-6.1 to 6.6). The respective correlation plots are shown in Fig.-6.1 to 6.6. The calculated values of hydrogen yield obtained through these correlations are compared against the experimental yield of hydrogen for different biomass samples (Table-6.5).

6.3.1 For sugarcane bagasse

$$(H_2)_{yield} = 1.4906 \times \left[(T)^{0.42} \left(\frac{S}{B} \right)^{0.07} (E.R.)^{-0.39} (\rho_M)^{-0.198} \right] \quad (6.1)$$

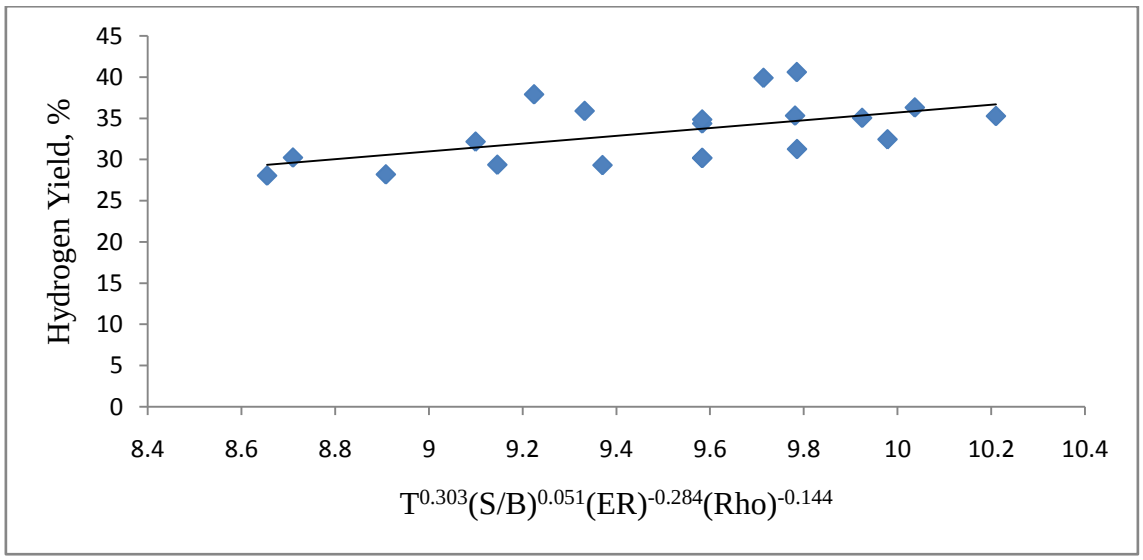


Fig. - 6.1 Correlation plot for yield of hydrogen against system parameters for sugarcane bagasse

6.3.2 for coconut coir

$$(H_2)_{yield} = 2.9494 \times \left[(T)^{0.3835} \left(\frac{S}{B} \right)^{0.23} (E.R.)^{-0.15} (\rho_M)^{-0.31} \right] \quad (6.2)$$

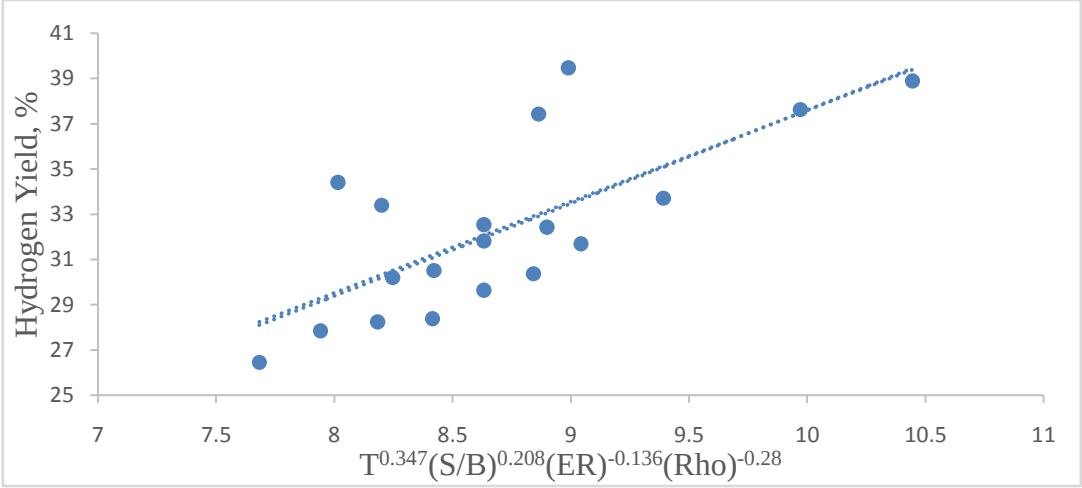


Fig.-6.2 Correlation plot for yield of hydrogen against system parameters for coconut coir

6.3.2 For rice Husk

$$(H_2)_{yield} = 1.3687 \times \left[(T)^{0.5465} \left(\frac{S}{B} \right)^{0.1157} (E.R.)^{0.09} (\rho_M)^{-0.1594} \right] \quad (6.3)$$

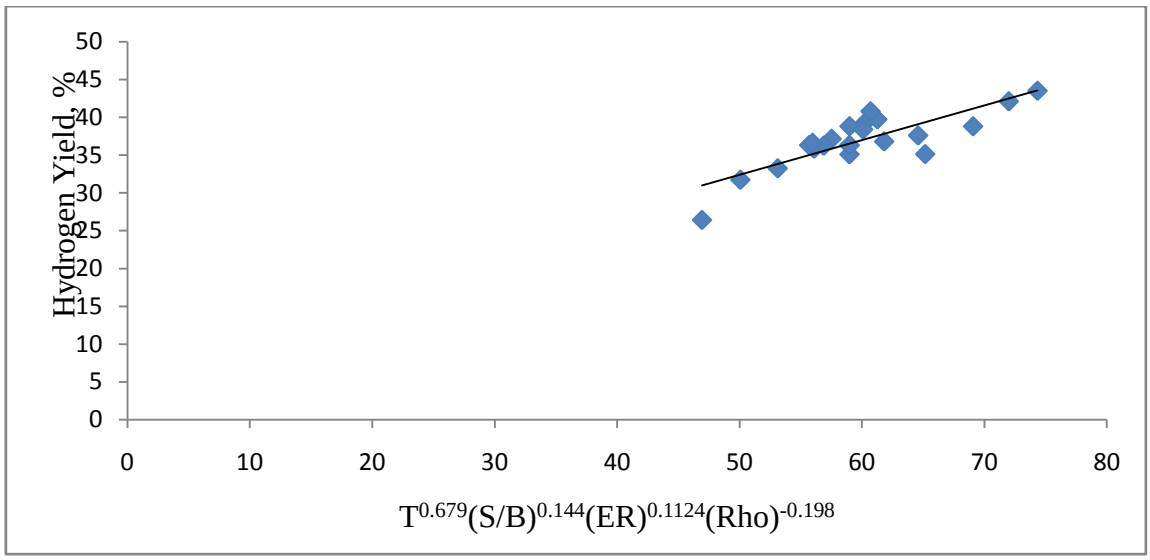


Fig.-6.3 Correlation plot for yield of hydrogen against system parameters for rice Husk

6.3.4 For wood Chips

$$(H_2)_{yield} = 0.3024 \times \left[(T)^{0.767} \left(\frac{S}{B} \right)^{0.036} (E.R.)^{0.224} (\rho_M)^{-0.028} \right] \quad (6.4)$$

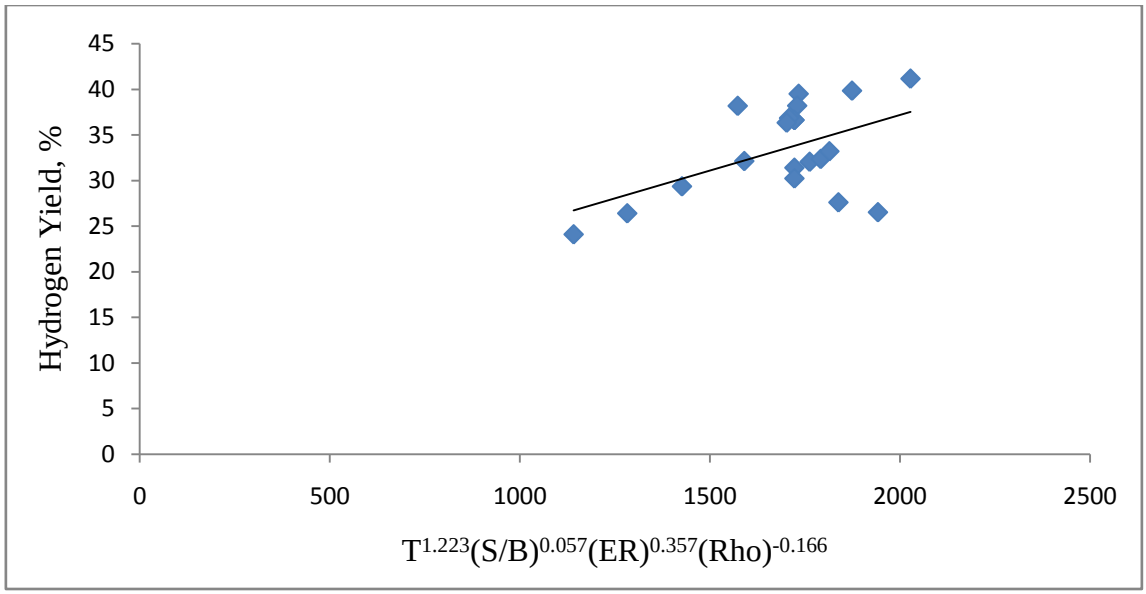


Fig.-6.4 Correlation plot for yield of hydrogen against system parameters for wood Chips

6.3.5 For rice straw

$$(H_2)_{yield} = 0.3217 \times \left[(T)^{1.118} \left(\frac{S}{B} \right)^{0.241} (E.R.)^{0.1985} (\rho_M)^{-0.147} \right] \quad (6.5)$$

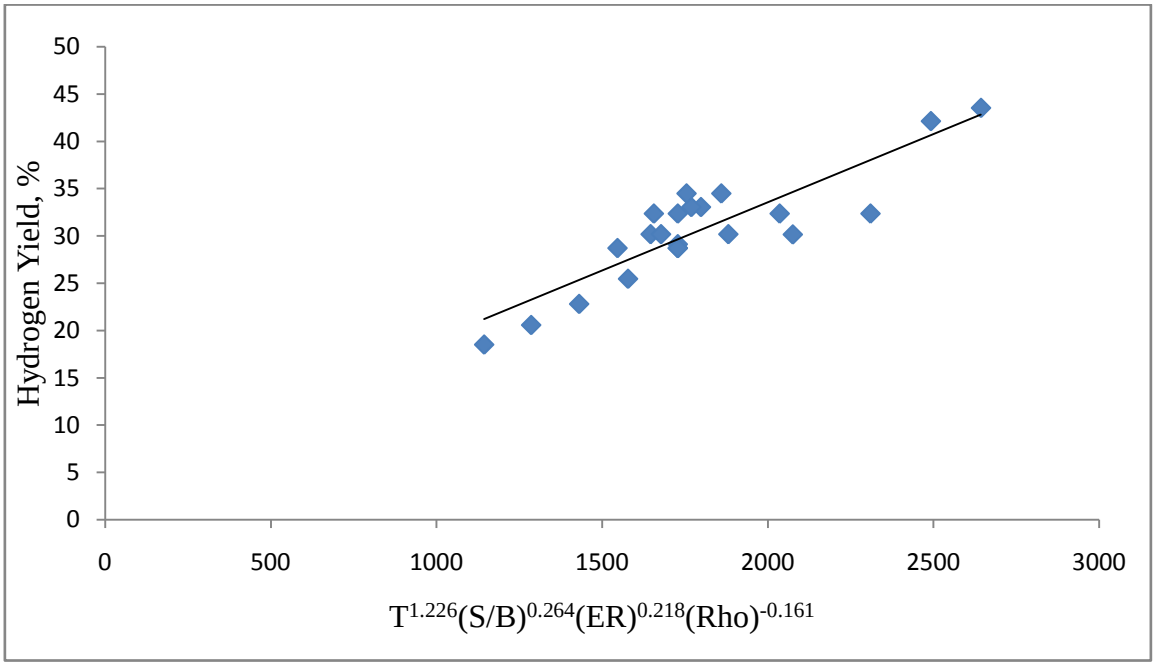


Fig.-4.5 Correlation plot for yield of hydrogen against system parameters for rice straw

6.3.6 For saw dust

$$(H_2)_{yield} = 0.0154 \times \left[(T)^{1.249} \left(\frac{S}{B} \right)^{0.2127} (E.R.)^{0.267} (\rho_M)^{-0.089} \right] \quad (6.6)$$

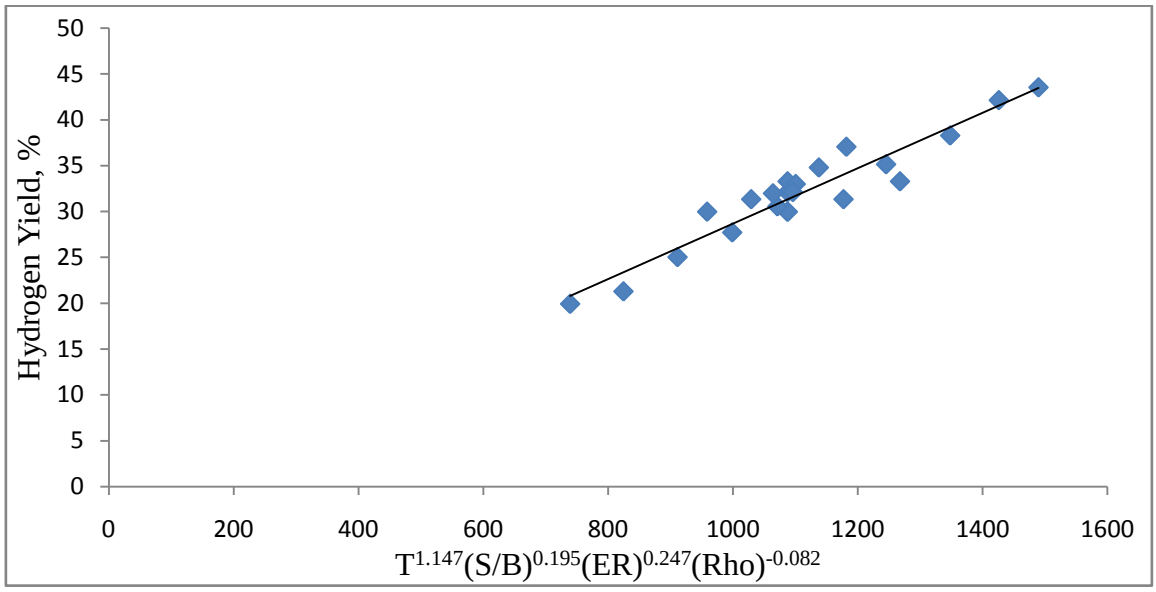


Fig.-6.6 Correlation plot for yield of hydrogen through gasification of saw dust

Table- 6.5: Comparison of calculated values of hydrogen yield against the experimental values

Sl. No.	Biomass Sample	Standard deviation, %	Mean deviation, %
1	Rice husk	5.83	-0.14
2	Rice straw	0.15	0.01
3	Saw dust	5.53	-0.13
4	Wood chips	13.71	-0.85
5	Sugarcane bagasse	8.98	-0.39
6	Coconut coir	7.70	0.30

The results obtained with respect to gasification process i.e. the hydrogen yield and gasifier performance are discussed here in this section. Before discussing the results it is essential to know the mechanisms of gasification of which the distribution of temperature in different zones is a part.

6.4 Different Zones of Gasifier

It is observed from Fig.4.10 that in the drying zone, temperature remains in the range of 150–300^{o}C during the time period of60 minutes. This implies that moisture content present in the biomass sample gets removed and temperature of the material increases. In the pyrolysis zone, temperature is about 600^{o}C where biomass samples are decomposed chemically

resulting in the production of char, tar, various oils and gases. In the combustion zone, oxidization takes place with the carbon conversion where temperature of the pyrolysis products provides the required heat. As a result gas-solid and gas-gas reactions take place causing the gasification of biomass samples to proceed. Heats released by the exothermic reactions are utilized for endothermic reactions. In the reduction zones, secondary reactions of biomass pyrolysis take place, i.e. cracking, reforming and tar decomposition. Nearly all these reactions are endothermic. Therefore the temperature of reduction zone drops from 800^oC to about 600^oC in this zone.

6.5 OnHydrogen Yield

The final composition of the product gas is determined by the amount of air and steam supply to the gasifier as well as the temperature and residence time of feed sample within the gasifier. The fractions of CO_2, CO, H_2, CH_4 and H_2O in the product gas are determined in the gasification process. Several variables affect these fractions. Some of these variables are discussed in this section.

6.5.1 Effect of Temperature

From Fig.-4.11 and 4.12, it is seen that increase in temperature from 500 to 1000^oC [Table-A-2, (Appendix-I)] H_2 content in syngas increases while CH_4, CO_2 and CO decrease. This may be due to the fact that water-gas shift reaction and water gas reaction are favored by high temperature which results in more hydrogen production. Again methanation is favored by low temperature resulting decreased methane yield with increased temperature. During combustion CO_2, H_2O and CO are produced but with further increase in temperature CO_2reactswithCto form CO, water-gas and water-gas shift reactions take place as a result CH_4, CO_2 and COcontents are decreased with increase in temperature. When comparison is made among different biomass samples (Fig-4.12),it is observed that under similar operating conditions, more H_2 and CO_2 are formed for wood chips, while coconut coir yields more CH_4 and sugarcane bagasse yields more CO in the product gas. The reason may be the presence of higher carbon content in wood chips as observed from ultimate analysis of the sample (Table -4.2) than other samples. The higher carbon contents oxidize to form CO_2, H_2O and CO which led to formation of more hydrogen. Proximate analysis reveals minimum ash content (Table-4.3) for coconut coir which might be the reason for more methane formation in the

product gas than other samples. Similarly low moisture content of sugarcane bagasse (Table-4.3) might be the reason for formation of more CO and $CO_2$with it than other samples.

6.5.2 Effect of Steam to Biomass Ratio

When steam to biomass ratio (S/B) is varied within 2.5 it is observed from Fig.-4.13 that only H_2 content increases and other components of syngas decrease with increase in S/B for all the biomass samples. Again it is observed that at any S/B ratio, coconut coir gives more H_2 and CH_4whichmay be due to low ash content of the sample than others. However more CO is yielded by wood chips and rice straw gives more CO_2. Above S/B ratio of 2.5, hydrogen yield decreases. This may be due to the fact that excessive steam might be lowering the reaction temperature thereby affecting the gasification process.

6.5.3 Effect of Equivalence Ratio

In Fig.-4.14, effect of equivalence ratio on yield of syngas has been compared component wise for different biomass samples. It is observed that up to ER of 0.25, H_2 increases with ER, and then decreases for all samples. The reason is that a small amount of air (not more than 25% of the stoichiometric amount for complete combustion) is required partial oxidation [5]. Other components of syngas are observed to decrease with increase in ER values. Reason has already discussed in sec 6.5.1 that when hydrogen yield increases yields of other components decrease. Rice husk is found to yield more H_2 while coconut coir is found to yield more CH_4. When ER is varied from 0.25 to 0.35, oxidation reaction becomes more significant than steam gasification reaction because of the presence of more oxygen (Table -4.2) and fixed carbon content (Table -4.3) with the rice husk sample. Again low ash content of coconut coir might be the reason for yielding more methane among all the samples with increased ER. A very small ER lowers the reaction temperature for which it is not favorable for biomass gasification. A large ER too decreases the retards the gasification reaction thereby affecting the hydrogen yield.

6.5.4 Catalytic Effects

Catalytic effects of bed materials on yield of syngas have been studied for different biomass samples by varying the bed materials. These effects for six different biomass samples are shown in Fig. 4.15 to 4.20 and respective data are listed in Table B-1 to B-6 in Appendix-I.

Each plot and table of one bed material has five sub-parts for different bed materials (i.e. only sand, 1:1 dolomite -sand mixture, 1:2 dolomite -sand mixture, 1:1 red mud-sand mixture and 1:2 red mud-sand mixture). It is observed that with the use of 1:1 and 1:2 red mud - sand mixture as bed material, yield of H_2 is found to be increasing with increase in temperature for all the biomass samples. This may be due to the fact that components of red mud might be giving better catalytic effect at higher temperatures (Figures - 4.15 and 4.20).

6.5.5 Overall Effects of Parameters

Chemical formula of all the feed samples are calculated and listed in Table-6.1 which gives the basis of calculation for stoichiometric air requirement for complete combustion. It is found that sugarcane bagasse contains more hydrogen than other biomass sample implying the possibility of yielding more energy.

Yield of hydrogen in kg/kg of feed sample, carbon conversion efficiency, cold gas efficiency, thermal conversion efficiency and gasifier efficiency with the energy equivalent amount of syngas to be produced per kg of feed sample are calculated through material and energy balances. Mass balance and energy balance calculations for sugarcane bagasse is shown here in this chapter while for five other samples these calculations are shown in Appendix-II. All these results along with the flue gas yield for different biomass samples are listed in Table-6.2 and Table-6.3. From this table it is observed that net energy output, thermal conversion efficiency and gasifier efficiency are more for rice husk while cold gas efficiency and carbon conversion efficiency are found to be more for rice straw. Reason for these may be higher void fractions with rice straw sample than that with rice husk. Because of compactness of the bed with rice husk, thermal conversion efficiency and net energy output might have been more than other biomass samples. Being less compact or more porous bed, all the rice straw might have combusted completely during gasification resulting in higher carbon conversion efficiency than other biomass samples.

Calculated values of hydrogen yield obtained through developed correlations (Eq. 6.1 to 6.6) are compared with the experimental values of hydrogen yield obtained through mass balance calculations. The comparison is shown in Table-6.5. It is observed that there is very good agreement for each of the biomass sample while standard deviation and mean deviation for

rice straw are almost negligible which indicates the occurrence of proper gasification.It is also observed from Eq.no. 6.1 to 6.6 (Fig. 6.1 to 6.6) that temperature is the significant parameter for yield of hydrogen in each case. Among S/B ratio and Equivalence Ratio, S/B ratio is found to affect the yield of hydrogen more. Effect of bed material on yield of hydrogen is found to be negative for all the biomass samples. On comparison it is seen that the effect of bed material is least for wood chips and saw dust.

Gasifier performance has been compared in terms of different efficiencies for different biomass samples in Table-6.4. It is seen that gasifier performance for rice husk is better as the gasifier efficiency is seen more with least deviation against the experimentally observed data than other samples. This may be due to the reason that operating conditions of gasification might have been suitable for rice husk. But cold gas efficiency for the saw dust sample is coming higher than that for other samples which may be due to the lower heating value of product gas as cold gas efficiency is the ratio of total energy output to energy input (i.e. heating value of product gas). On comparing gasifier efficiency against the cold gas efficiency minimum deviation is observed with the rice husk sample. More specifically it can be said that gasifier performance is observed to be very good with rice husk sample with minimum deviation and maximum net energy output among all the biomass samples

(B) CFD Simulation Results and Discussions

CFD simulation results with respect to hydrodynamic studies and thermal flow behavior with no reactions are discussed in this section.

6.6 Contours of Solid Volume Fraction

The contour plots of volume fraction of sugarcane bagasse at air inlet velocity of 0.07m/s are shown in Fig 6.7. Simulation was carried out at different air velocities Fig. A (a-d) in Appendix-III. It is observed that at 0.9m/s air velocity better fluidization is achieved. Therefore 2D simulation was carried out for six different biomass samples at 0.9m/s air velocity and contour plots of solid volume fractions are shown in Fig. BI (i-vi) in Appendix-III. Similarly 3D simulation contour plots of solid volume fractions for six biomass samples are shown in Appendix-III [Fig.-BII(i-vi)]. Contour plots of air volume fraction for both 2D and 3D simulations are shown in Fig. CI and CII while same for bed materials are shown in Fig. DI and

DII respectively. It is observed from Fig.6.7that the bed begins to expand at this velocity of 0.7m/s and the particles start moving slightly. But only with a small increase in air flow the particles just start to exhibit fluid-like behavior at 0.9m/s [Fig. A(d) in Appendix-III]. Thus the upward drag on the solid material becomes equal to the weight of the material at this condition.

The contour plots of sugarcane bagasse, air and sand volume fractions for 2D and 3D simulations at 0.9m/s velocity and time of 40sec with the static bed height of 10cm are compared in Fig.6.8. It is observed from Fig.6.8 that bubbles are formed only within the static bed height without any noticeable bed expansion. The reason may be attributed to the fact that bubbling occurs at the surface only. In other words, solids in the bottom section of the bed are in pneumatic transport while fluidization in the upper section is in freely bubbling state.

From Fig.6.7 it is also seen that the variation in the bed profile with time for sugarcane bagasse at air velocity of 0.7m/s is very clear and the bed profile changes with time. But after some time no significant change is observed in the bed profile. This indicates that the fluidized bed has come to a quasi-steady state. The 3D-contour plot in Fig.6.8 shows higher solid volume fractions along the walls are more in comparison to the core region. This may be due to the segregated tendencies of the particles towards the walls or gulf streaming. Thus the solid particles slide down along the wall of the reactor without much resistance and move towards core region to be picked up by the gas again. From Fig.6.9(A), it is observed that the contours of volume fractions of sugarcane bagasse, sand and air obtained at air velocity of 0.9 m/s for initial static bed height 0.1m in 2-D/3-D fluidized bed have achieved the quasi steady state. The contours for sugarcane bagasse and sand illustrate that bed is in fluidized condition. The contour for air illustrates that volume fraction of the gas is less in fluidized section than the solid particles.

The solid volume fractions for different biomass samples in the fluidized bed at air velocity of 0.9m/s are compared with each other [Fig.6.9 (B)]. It is observed from this figure that solid volume fraction is at about 0.325m bed height (0.3 – 0.35m) for all samples indicating the proper fluidization condition where solid volume fraction is maximum in the top region of the bed. Then it decreases to zero in the upper part of the fluidization column where gas is only found.

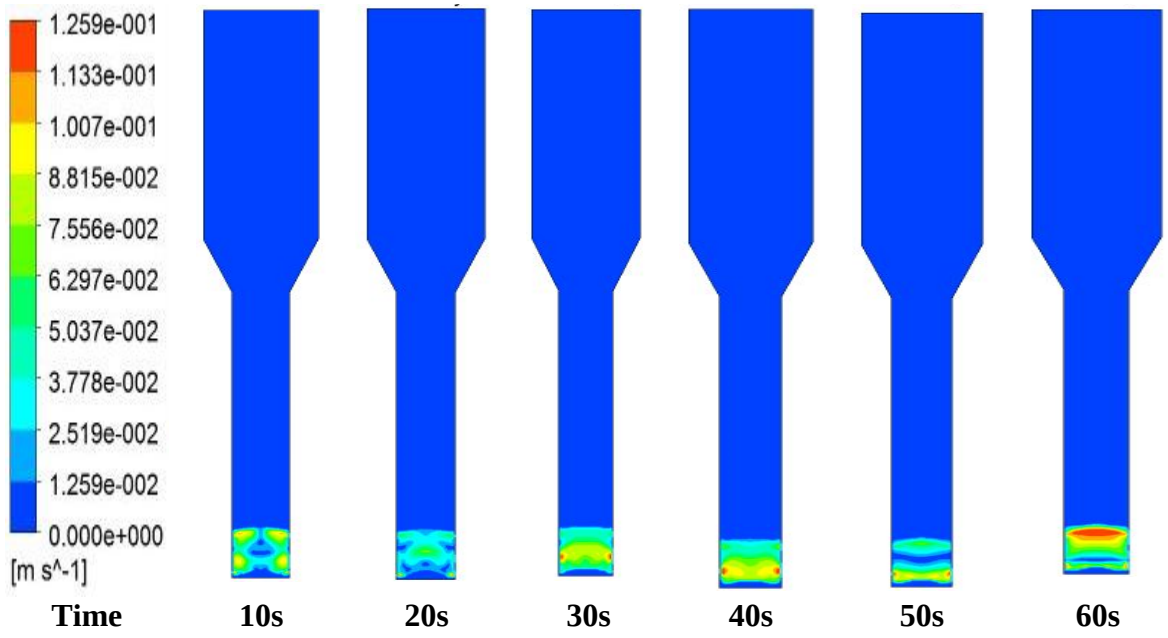


Fig. 6.7: Contour plot of volume fraction against time for sugarcane bagasse at air velocity of 0.7m/s for initial static bed height of 0.1m

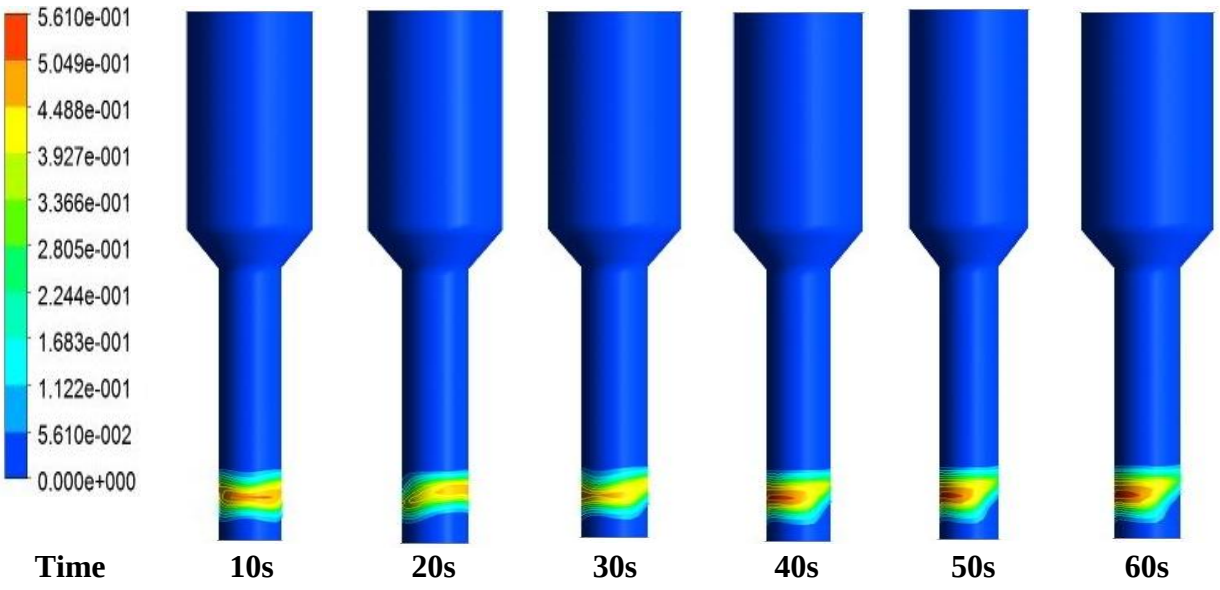
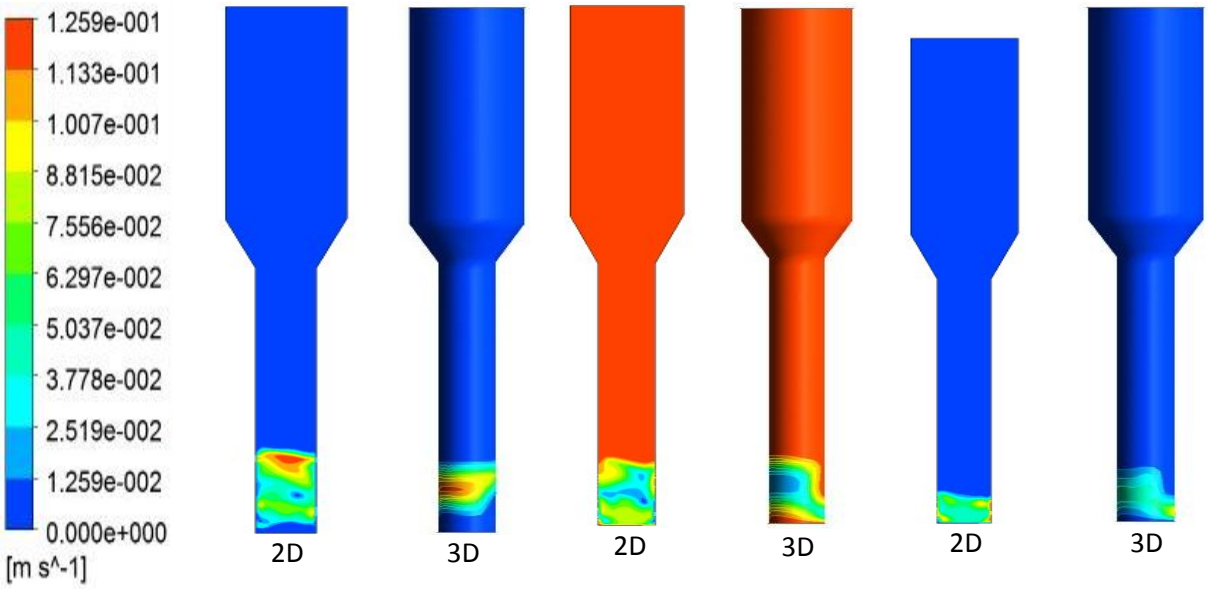


Fig. 6.8: 3D - Contour plot of volume fraction of sugarcane bagasse at air velocity of 0.9m/sec with respect of time for initial static bed height of 0.1m



Time | Sugarcane bagasse volume fraction | Air volume fraction | Sand volume fraction

Fig.-6.9 (A) Comparison of 2D -3D contour plots of volume fractions for sugarcane bagasse, air and sand at 0.9m/s velocity, 40sec time and initial bed height of 0.1m.

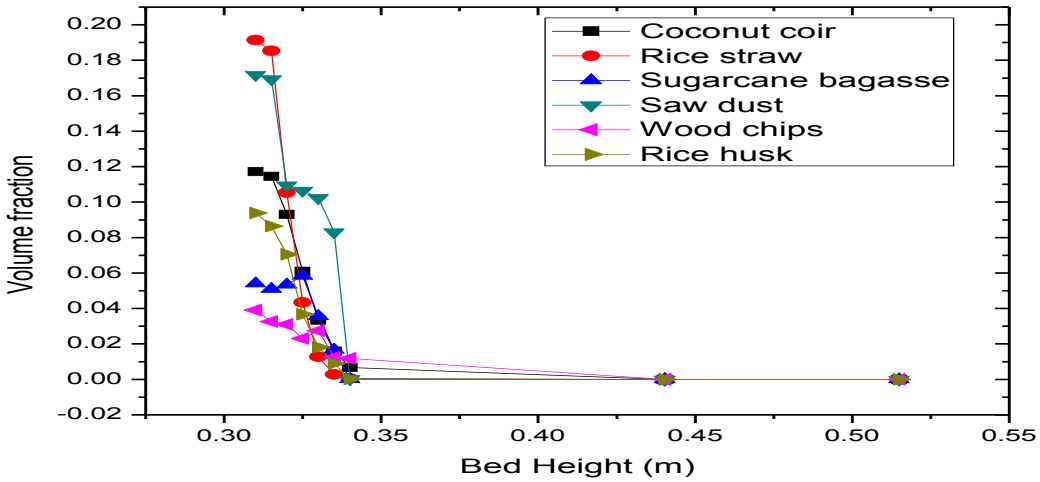


Fig. – 6.9 (B): Comparison of Solid Volume fractions for different biomass samples in the fluidized bed at air velocity of 0.9m/s.

6.7 Phase Velocity

The velocity vectors show magnitude of velocity with direction and thus helpful to determine the flow pattern in fluidized bed. The velocity vectors of sugarcane bagasse, sand and air in the column obtained after the quasi steady state at air velocity of 0.9 m/s with initial static bed height of 0.1 m are shown in Fig.6.10 and 6.11.From velocity vector of solid phase (Fig.-

6.10), it is observed that there is vigorous movement of solid particles throughout the bed implying that the velocity at the bottom is less. In the central region of the bed, direction of velocity near the wall is observed to be downwards while that in the region away from wall is upwards. In the upper part of fluidizing section there is circulatory motion/ downward motion of the solid particles near the wall and upward motion in the central region of the bed. The velocity vector of gas phase in the column (Fig.-6.11) indicates that there is an upward flow throughout the column which implies that velocity of air is very less within the bed compared to that in remaining part of the column. This is due to very small volume fraction of air compared to that of solids within that particular region of the bed. In the upper section of the column, air velocity is high as it carries air bubbles but in the lower section of the column solid particles obstruct the movement of bubbles which reduces air velocity. Vector plots for different biomass samples are listed in Fig –EI (i-vi) and EII (i-vi) in Appendix-III for 2D and 3D simulations respectively. Comparison of axial velocity profiles for six different biomass samples are shown in Fig.6.12. It is seen from this figure that almost all the particles attain velocity of 0.45 to 0.55 m/s in the bed height zone of 0.25 to 0.35m indicating their proper fluidized state. It is further seen that at 0.35m of bed height some particles attain maximum velocity of 0.9m/s which may be due to the fact that light/fine particles rise up with the supplied air flow and attain higher bed heights. During fluidization more fine particles are produced because of inter particle and wall-particle collisions. As a result more particles go to the freeboard region where impact of air flow reduces. Therefore velocity of particles gets decreased but these particles move up further up to bed height of 0.55m by inertia force and collision force.

6.8 Bed Pressure drop

The axial pressure drop in a fluidized bed varies from higher value at the bottom of the bed to zero value at the top of the column. The bed pressure drop can be determined from the difference of pressure at the inlet and outlet. Fig.6.13 shows the contours of statics gauge pressure. It is evident from this figure that the pressure is higher at the inlet and gradually decreases and became zero at the outlet. Pressure drop contour plot profiles for all the biomass samples by 2D and 3D simulations are shown in Fig.-FI and FII (Appendix-III) respectively. Pressure drop profiles at air velocity of 0.9m/s for different biomass samples are

compared in Fig.-6.14. It is seen that all the biomass samples show almost constant bed pressure drop implying all the biomass samples are in proper fluidization condition. This has already been observed from effect of velocity that all the biomass is getting fluidized properly at 0.9m/s. As during fluidization all the bed materials attain fluid state and gravitational force is just balanced by drag force as a result the bed pressure becomes constant. This is seen in above figure (Fig. 6.14). It is further seen that magnitude of pressure drop is minimum for rice straw and wood chips (with more void spaces) while it is maximum for saw dust and coconut coir (with less void space). Reason for this may be the void space. Structure of rice straw and wood chips create more voids in the bed while less void space is created with saw dust and coconut coir.

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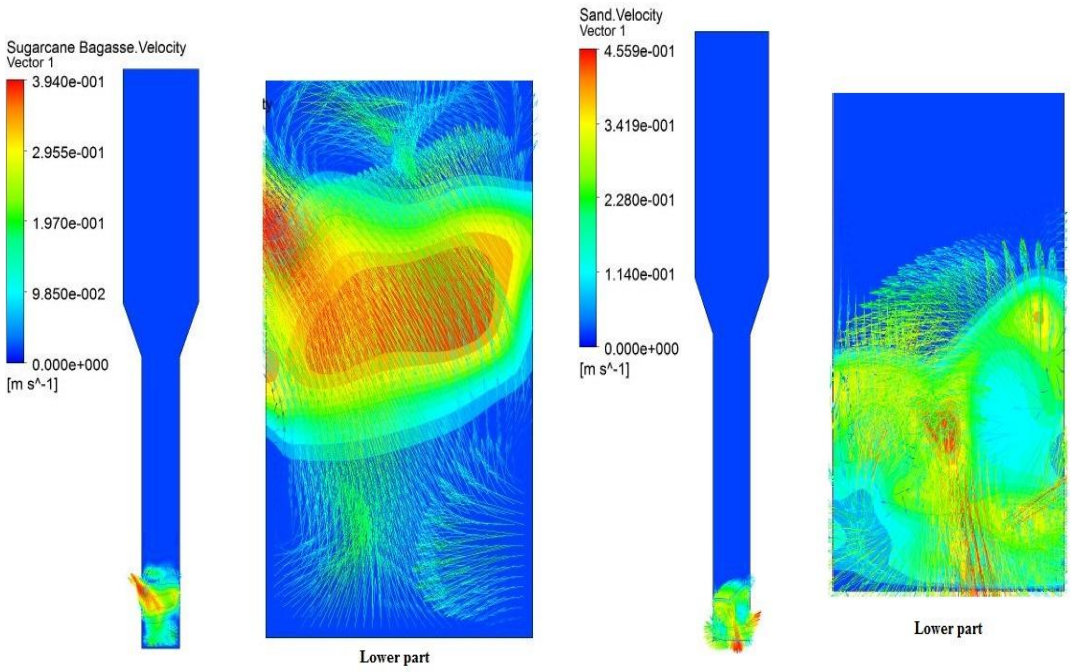

Fig. 6.10: Velocity vector of sugarcane bagasse and sand at air velocity 0.9 m/s

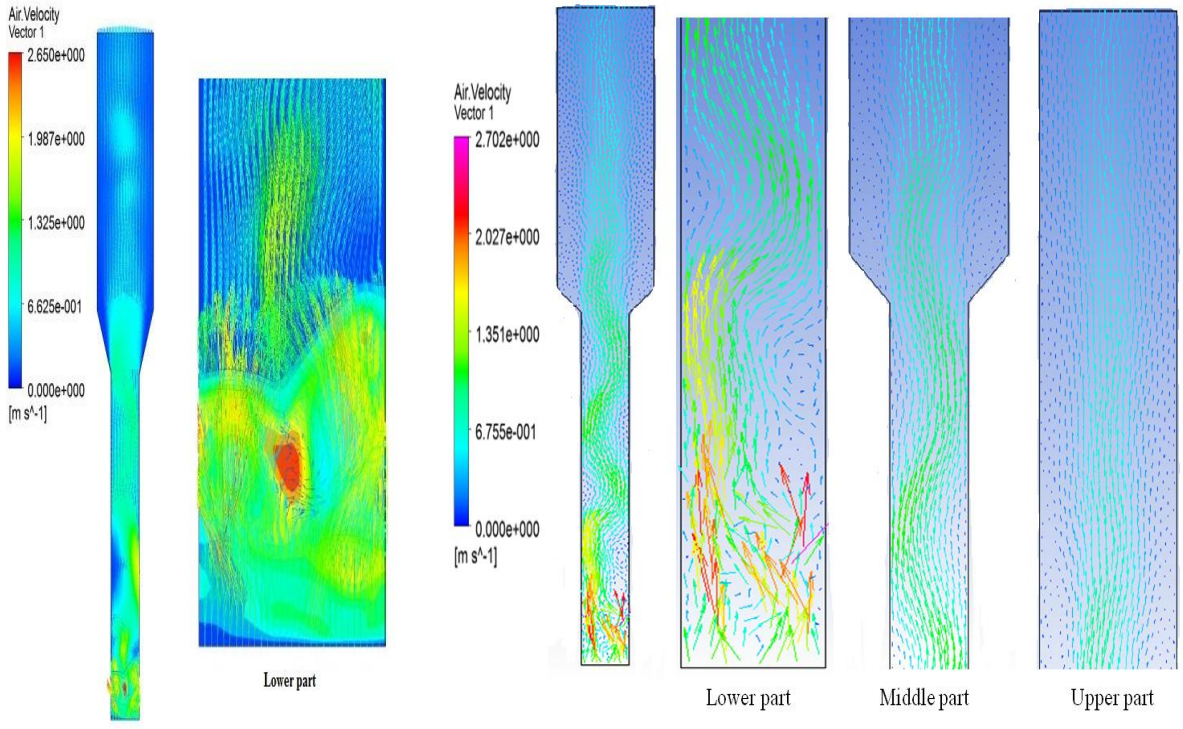

Fig. 6.11: Velocity contour and vector plot of air in different parts of the fluidized bed

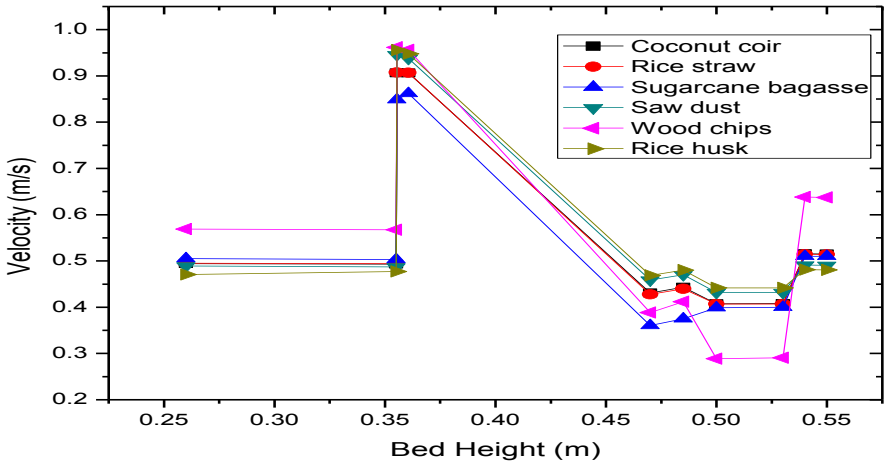

Fig.-6.12- Comparison of axial solid velocity profiles in the fluidized bed for six different biomass samples

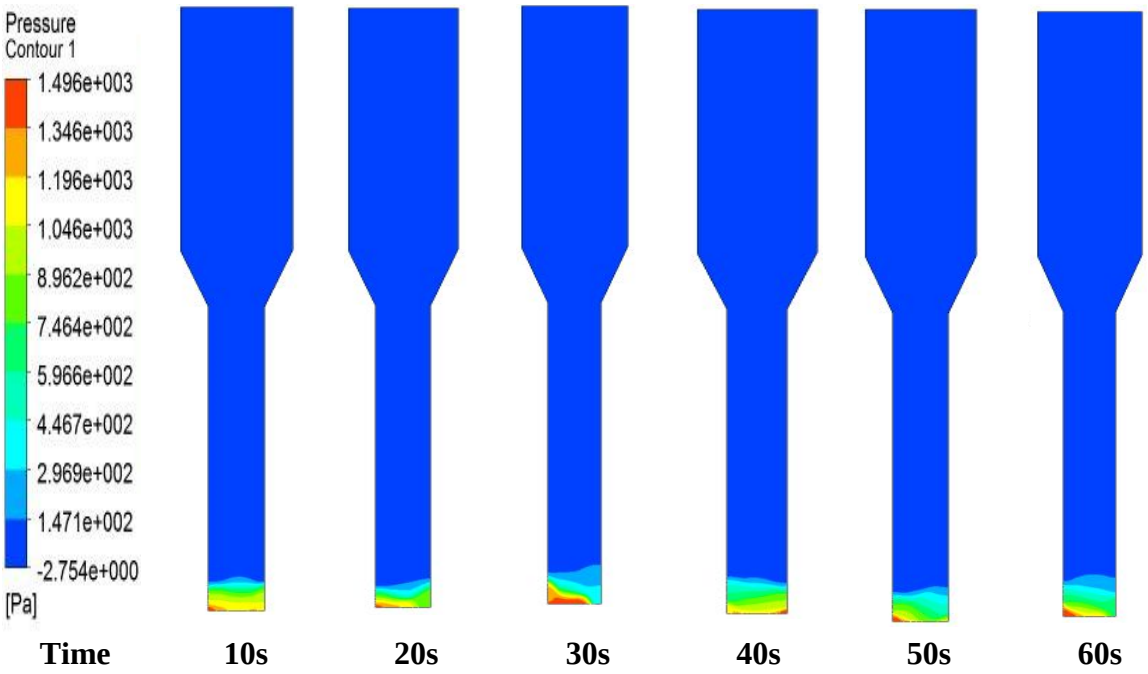

Fig.-6.13: Contour plot of bed pressure drop for sugarcane bagasse at air velocity of 0.9m/s

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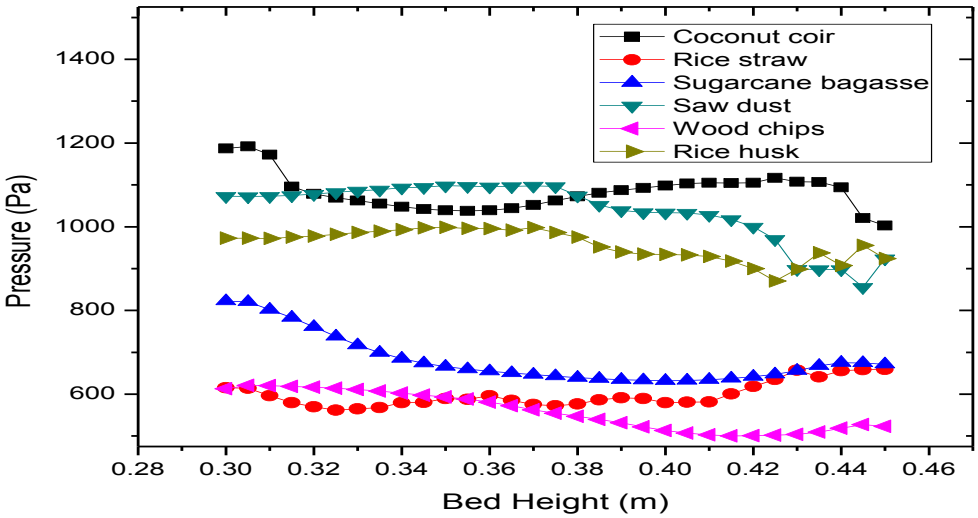

Fig. 6.14: Comparison of pressure drop profiles in the fluidized bed at 0.9m/s air velocity for different biomass samples.

6.9 Effects of Inlet Velocities

The volume fraction distribution for the particles using the Gidaspow model with four inlet velocities, i.e., 0.7 m/s, 1.0 m/s, 1.8 m/s and 2.0 m/s, are shown in Fig. 6.15(a-d) for particles. If the gas velocity does not exceed terminal velocity of particle, V_t the particles fall back to the bed. This is referred to as a bubbling bed and is shown in Fig. 6.15 (a)& (b). Exceeding V_t means the suspended particles are carried up with the gas phase and continue to move in the riser and fluidize in the freeboard area resulting in fast fluidization. This state has been shown in Fig. 6.15 (c)& (d). The contour plots of Fig.6.15 (a)& (b) show bubbles increasing in size and distorting with increasing height.

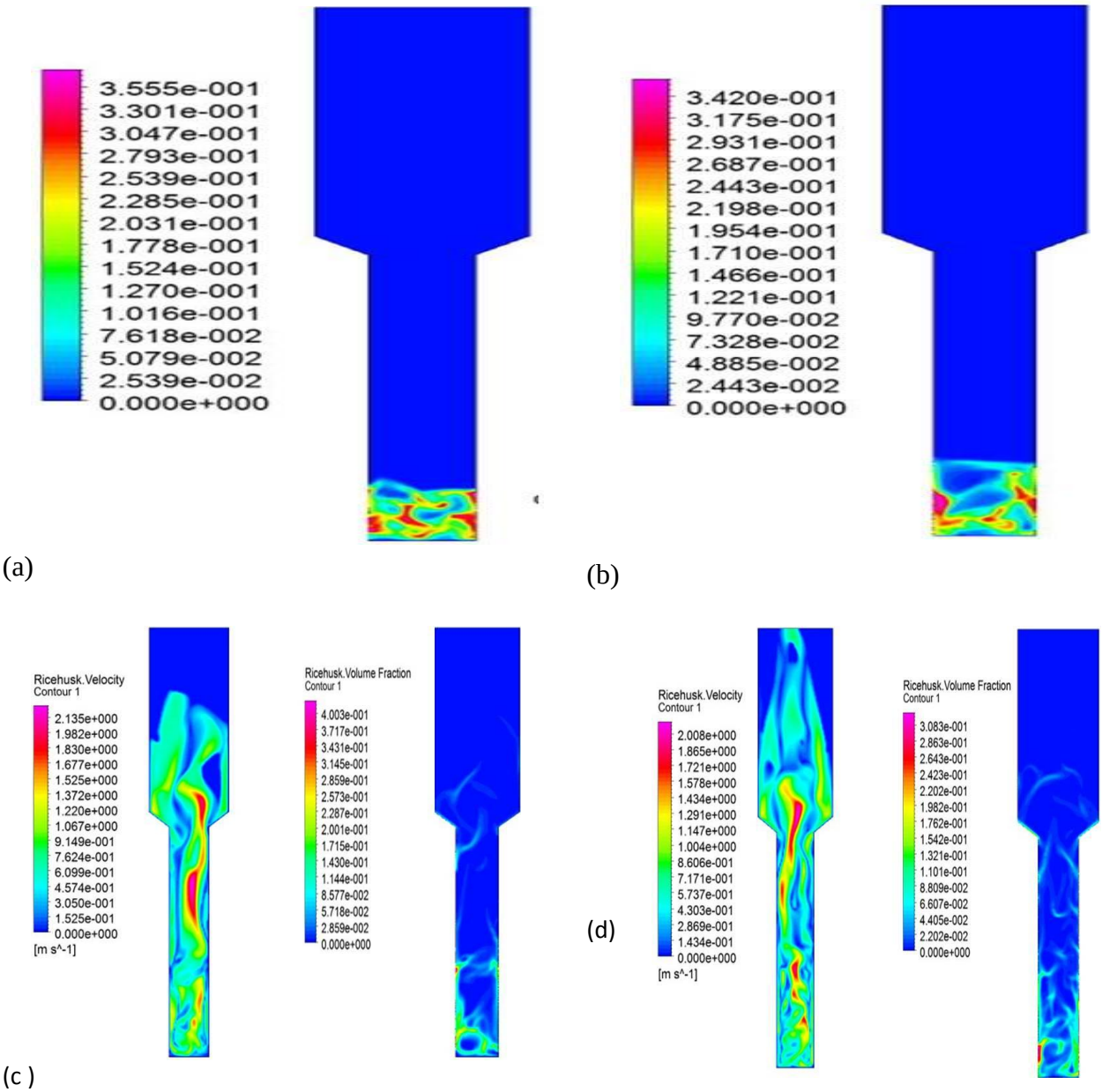

Fig. 6.15: Particle volume fraction and velocity vector for biomass sample with d_p = 530 µm:
a) V = 0.7 m/s, b) V = 1 m/s, c) V = 1.8 m/s d) V = 02 m/s.

This is due to the coalescence of the bubbles with smaller upcoming bubbles known as trailer bubbles. As the velocity increases, the bubble sizes increase and the solid-gas mixture becomes more dilute particularly towards the top of the bed. The solids are pushed down by the bubbles, and then they descend to the base of the reactor.

The gas – solid phase in fast fluidizing state in Fig. 6.15 (c) & (d) appears to be very dilute in comparison to the bubbling models. The particle volume fraction and particle velocity are

shown in Fig. 6.15 (c) at1.8m/s gas velocity which is slightly lower than the terminal velocity. Increase in gas velocity allows for a faster flow of gas to push the particles upward in the bed.Fig.-6.15 (d) shows the particle volume fraction and particle velocity at gas velocity 2m/s where terminal velocity is just exceeded (terminal velocity= 1.9m/sin this case).

6.10 Thermal-Flow Behaviour with no Reactions

Thermal-flow behavior of the particles is studied in the fluidizer. Sand and biomass sample particles are patched up to a static bed height 0.1m.The air enters at a velocity of 0.9 m/s at 673 K and flows through the bed.It isseen that air streams move upward whereas particles circulate within the fluidized bed in the bottom part of the domain. At 0.9 m/s inlet air velocity, no particles are seen in the upper part of the domain. A sequence of volume fractions distributions for biomass is already shown in Fig.6.7 at different times. Bubbles are formed at the distributor (entry point of air to the fluidizer) due to the continuous supply of air at a rate of 0.9 m/s. The bubbles continue to rise up till it breaks or reaches the top of the bed by making its path by pushing the particles sidewise. Bubbles also grow in size during its movement by coalescing with other bubbles. Solid particles are dragged upward by the bubbles as wake solids. When the bubbles break the solid particles are thrown towards the wall. The solid particles are also pushed towards the wall region by the bubbles during its movement in upward direction. When the solid particles reach the wall region they slide down along the wall, move side wise to fill the void spaces created by movement of bubbles. Thus the circulation of solid particles is always there in the fluidized bed by the movement of bubbles. This indicates the solid particles in the bed move in a circular motion thereby influencing and distorting the bubbles. As a result bubbles tend to adhere to the wall and move along the wall. This is more clearly evident in Fig. 6.10 which displays the particle velocity vectors. Since no reactions are simulated in this case, the temperature inside the domain is considered to be uniform. A sample plot for sugarcane bagasse is shown below in Fig. -6.16. Temperature profiles for other samples with 2D and 3D simulations are shown in Fig – G in Appendix-III. Solid temperature at different bed heights within the gasifier are compared for different biomass samples (Fig. 6.17). From Fig. 6.17 it is observed that almost uniform temperature is maintained within the gasifier for all the biomass samples in the zone of 0.25 to 0.50m bed heights. It is further seen that bed temperature decreases slightly beyond

0.5m bed height indicating the completion of gasification process and production of syngas. As already mentioned initial bed height of 0.1m is considered in the fluidized bed gasifier. By the time bed materials and biomass particles fluidize properly, bed height of 0.25m is attained. Drying and pyrolysis process also take place. During pyrolysis the biomass particles start decomposing chemically and particle size gets reduced by collision during fluidization. As a result the fine particles along with gas move up further. Oxidation and reduction take place leading to completion of gasification. As it is a continuous process by the time gasification gets completed for first batch of feed, drying, pyrolysis of second batch feed initiate. Thus for the entire period of gasification all the reactions go on continuing and flue

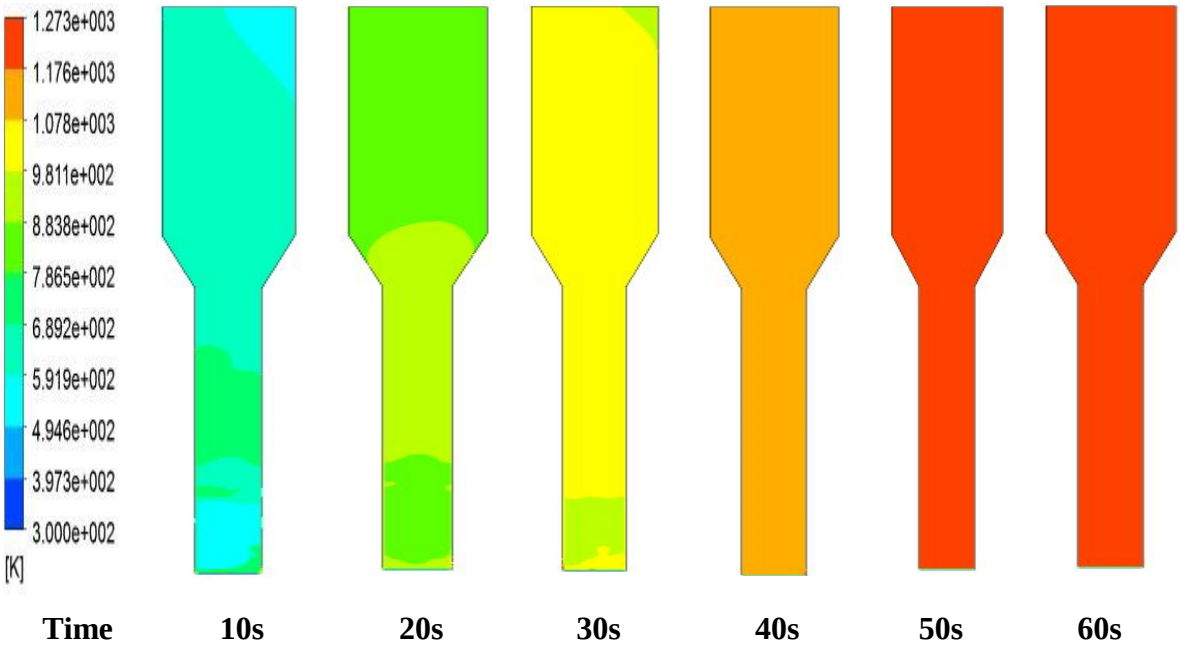

Fig. 6.16 Temperature profile at different times inside the fluidized bed for Sugarcane bagasse at 1273K gasification temperature.

gas gets produced from the gasification process. Once the flue gas starts producing, the bed temperature decreases. This is seen in Fig. – 6.17 where bed temperature decreases after 0.5m bed height. Comparing the temperature profiles for different biomass samples it is observed that temperature for rice husk and saw dust are less and that of rice straw and wood chips are more in the same zone of bed height. This may be due to the fact that rice husk and saw dust show lower void spaces while rice straw and wood chips exhibit more void spaces among the

particles. Again structure of rice husk and saw dust make them more compact for which temperature profile in the gasifier might have been affected.

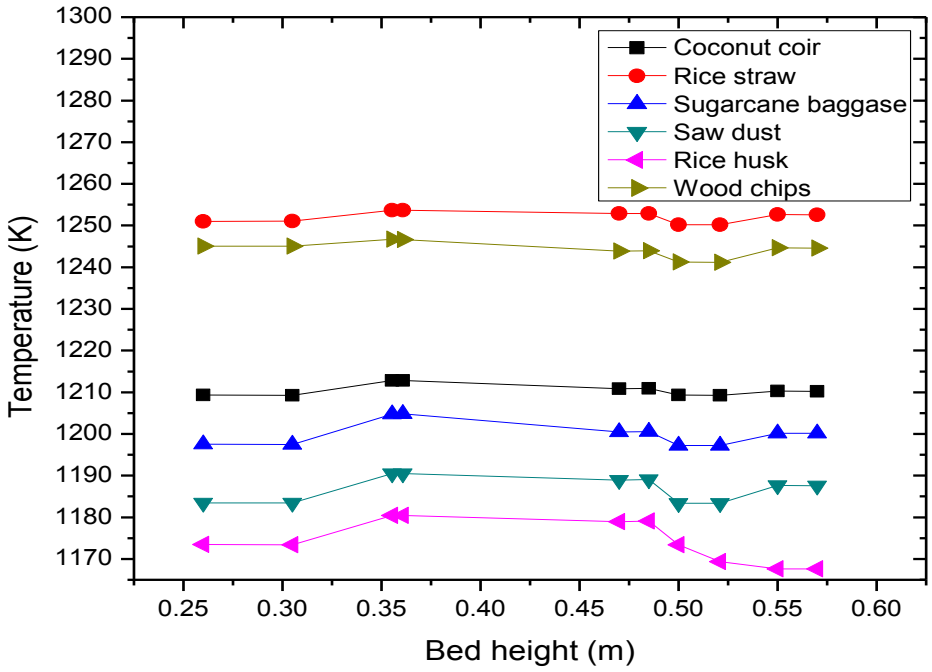

Fig.- 6.17 : Comparison of Temperature profiles against bed height within the gasifier for different biomass samples with ER=0.25, S/B=0 and T=1273K.

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Chapter-VII

Conclusions

An indigenous fluidized bed gasifier is designed and installed in the laboratory. Different biomass samples are gasified in this fluidized bed gasifier with different bed materials. Compositions of syngas are analyzed by a gas analyzer. Physical properties of the sample are determined before experiments. Proximate and ultimate analyses are also carried out for all the biomass samples. Using these results chemical formula for different biomass samples is determined. Based on the chemical formula stoichiometric air requirement for complete combustion of biomass samples are determined and equivalence ratio is varied accordingly. Energy balance and mass balance calculations are carried out to determine the net energy output and gasifier performance in terms of various efficiencies such as carbon conversion efficiency, cold gas efficiency and thermal efficiency The effect of different system parameters (viz. temperature, steam to biomass ratio (S/B), ER and bed material) on the yield of hydrogen in the syngas are studied. Calculated values of hydrogen yield are compared against the experimental yield of hydrogen for validations of developed correlations.CFD simulations are also carried out for the Fluidized bed gasifier. Thus the overall conclusion can be drawn in the following three aspects.

7.1 On Hydrogen Yield

The correlationsfor the yield of hydrogen are developed for different biomass samples on the basis of dimensional analysis by relating experimentally observed H_2-yield against different system parameters. From different experiments it is observed that optimum yield/conditions for more hydrogen yield are obtained at S/B=0.5, ER=0.25 and temperature=700^{o}C for all the biomass samples. Use of red mud-sand mixture as bed material is found to impart better catalytic effect on H_2 yield than dolomite-sand mixture. In most of the cases, 1:1 red-mud - sand mixture yield more than 1:2red-mud - sand mixtures implying that more amount of red mud in bed material provides better/suitable condition for production of hydrogen (H_2-yield).

Again calculated values of H_2-yield are compared against the experimentally observed data. It is observed that very good validation is there which impliesthat the developed correlations can be used over a broad range of parameters for all types of biomass samples. Thususing these correlations, approximate information on H_2-yield from different biomass samples can be obtained prior to the real experimentation. Otherwise it can be concluded that the developed correlations may be used as the basis of design for scale up of the gasifier toproduce more energy from biomass wastes.

Amount of hydrogen produced from different biomass samples are observed to vary within 56 to 74 gm per kg of feed sample. It is further observed from the present work that net energy produced per kg of feed sample for saw dustandrice husk are more than other samples. For rice husk and saw dust the yield of energy are 6.21 and 5.92kWh respectively. The energy yield fromwood chips is found to be minimum (5.09kWh) among six different samples.Although chemical formulas of these samples indicate that more hydrogen is present with sugarcane bagasse and then with coconut coir; yield of hydrogen is found to be more from rice husk. Wood chips ranks second in energy production among six biomass samples. Thus it can be concluded that sugarcane bagasse and coconut coir could not be gasified completely. The moisture content in these feed samples might be more than the required level. The reason for this may be the bigger size of particles in comparison with rice husk which might have caused improper fluidization for which drying might have affected. Rice husk might have undergone complete conversion during gasification due to permissible level moisture content and proper fluidization.

The energy reports on these samples reveal that the gasification process is cost effective. The energy results obtained through these experiments are found to be very much satisfactory. Thus the process can be scaled up for industrial uses.

7.2 On Gasifier Performance

Experiments are carried out in the in-house fluidized bed gasifier. Production of syngas validates the design. Operating conditions are varied to know the optimum conditions. The experimental observations obtained with the optimum conditions are considered for energy and mass balances as well as for the performance studies of the gasifier. Gasifier performances are determined for each biomass sample in terms various efficiencies vizcarbon

conversion, cold gas, thermal conversion and gasifier efficiency. It is observed that gasifier efficiency is deviating only within4% from the thermal conversion efficiency which implies that the performance of the gasifier is very much satisfactory. Again Carbon conversion efficiency is found to vary within 70 to 97% which indicates that gasification has been carried out properly and results obtained are satisfactory. Thus it can be recommended that temperature of 700^{o}C, ER of 0.25, S/B ratio of 0.5 and 1:1 red mud-sand mixture as bed material arethe optimum process parameters for this fluidized bed gasifier. Satisfactory amount of H_2-yield is observed with each biomass sample and the process is found to be very much cost effective through energy audit report Therefore it can be concluded that gasification of biomass samples using fluidized biomass gasifier is ecofriendly and can be very much profitable.

7.3 On CFD Simulation

The comprehensive information concerning the thermal-flow behaviour and gasification process existing inside the fluidized-bed gasifier is obtained through CFD simulation in this study. Based on the results obtained in the CFD simulation, the following conclusions are drawn.

- Increasing superficial gas velocity makes the flow development faster implying that the superficial gas velocity has a strong influence on the axial solids velocity and subsequently on the down flow of solids.
- The bed expansion behavior is found to vary with variation in gas velocity.
- Model is able to describe quantitatively the accumulation of solid at the wall. Solid concentrations appear flat in the core and increase towards the wall region.
- Back-mixing behavior or accumulation of particles is found to exist in the fluidized bed.Through vector plots, the velocity in the core region are found to be in upward flow direction and much higher than that in the annulus region, while solid and gas velocities near wall are found to be decreasing.
- The CFD simulation exhibits a solid circulation pattern for all the operating conditions with all the biomass samples which is observed to be consistent with the literatures reported by various investigators.

- The velocities of the fine particles are found to be larger than that of the bigger particles in the lower zone due to the attainment of high slip on the bottom side. The volume fraction of big particles is lower in the upper region than that in the lower portion of the bed.
- The simulation models achieved in predicting the bed dynamics of the fluidized bed gasifier such as its temperature and pressure distribution are also found to be satisfactory.
- Multiple phases for the bed with biomass sample and sand, led to phase segregation as lower density biomass particles migrated towards the top of the bed and the denser sand particles remained in the bottom part of the reactor.
- Assumption of uniform temperature led to homogeneous reaction. Therefore distribution of species and gas temperature indicate that reactions to occur very fast and finish very quickly implying a 100% carbon conversion.

7.4 General Observations for the present work

Design of the gasifier is found to be very much satisfactory as in-house FBG with the design functioned properly. Gasification is a complex process. Before completion of preheating or drying, fluidization starts and combustion takes place. Some of the observations are as follows.

(i). Gasification process is carried out with four equivalence ratios. With these ratios more air might have passed to the gasifier. As a result noticeable amount of free oxygen are foundwith the flue gas as indicated by gas analyzer. Oxygen amount has also been found to be varying within 13-19% of total amount of flue gas for different biomass samples. Complete gasification of biomass samplemight have been affected due to more air and thus the hydrogen yield might have affected in reality.

(ii)Steam to biomass ratio was also varied only four times randomly from which any solid conclusion cannot be drawn. Thus these four values of S/B ratio might have caused improper gasification of biomass samples affecting the yield of hydrogen.

Thus it can be said that if the gasifier is designed properly or scaled up with these information, energy demand of the nation can be fulfilled in an ecofriendly manner by biomass gasification to some extent.

7.5 Future Scope of the Work

During biomass gasification it is observed that some char particles and un-burnt biomass samples are remaining within the gasifier. Process conditions are determined from chemical formula. But high temperature and residence time combination needs more attention for which it is felt that some work can be carried out on the char and un-burnt particles. As gasification is a complex process, more computational work is required to understand it thoroughly. During syngas production much of heat is lost to the atmosphere.

Therefore the further works which need some attention can be outlined as follows.

i. More experiments can be carried out with different temperature and residence time combinations for the complete conversion of the biomass samples.
ii. Analysis of char for some value aided end products viz. extraction of silica and activated carbon
iii. Design of storage unit for syngas
iv. Separation of Hydrogen from syngas
v. CFD simulation for complete gasification process
vi. Studies on characteristics of biomass which will yield more energy.

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Appendix -I

(Experimentally Observed Data)

A. Effect of different system Parameters on Syngas

Table-A-1 Variation of temperature inside the gasifier

Temperature Profile				
Time (min)	Drying Zone ^{0}C	Pyrolysis Zone ^{0}C	Oxidation Zone ^{0}C	Gasification and ReductionZone ^{0}C
0	155.07	159.16	716.91	795.36
5	151.07	165.74	746.80	778.93
10	157.60	197.59	793.45	777.00
15	181.07	216.77	882.34	779.27
20	187.69	235.94	950.05	781.52
25	190.02	227.69	937.21	777.49
30	232.55	255.33	851.93	754.74
35	241.11	278.62	836.28	754.81
40	296.35	369.66	769.14	752.98
45	334.61	411.95	762.92	753.05
50	385.59	534.64	765.21	740.64
55	385.88	602.56	768.25	736.84
60	386.12	602.95	768.26	737.02

Table- A-2 (a): Effect of temperature on Syn-gas composition for rice husk

(S/B= 0, Equivalence Ratio =0.25, Feed Rate =10kg/hr, Bed material= Silica Sand)

Temp(^{o}C)	H_2	CO_2	CH_4	CO	NHV kcal/m^3
500	21.5	30.3	10.7	37	2382.78
550	24.57	29.8	10.2	35	2403.76
600	26.8	28.17	9.56	34.65	2357.10
650	30.46	26.77	8.24	33.41	2295.84
700	36.21	24.4	7.16	31.32	2279.63
750	37.18	23.6	8.33	30.2	2380.71
800	38.36	24.3	8.11	28.7	2413.28
850	38.05	23.95	8.02	27.3	2387.05
900	38.12	23.88	7.98	27.01	2383.3132
950	39.49	23.89	7.28	28.09	2358.83
1000	40.58	22.45	7.37	28.91	2351.16

Table- A-2 (b): Effect of temperature on Syn-Gas composition for rice straw

(S/B= 0, Equivalence Ratio =0.25, Feed Rate =10kg/hr, Bed material= Silica Sand)

Temp(^oC)	H_2	CO_2	CH_4	CO	NHV Kcal/m^3
500	18.5	31.30	10.7	40	2335.82
550	20.57	30.80	10.2	38	2331.10
600	22.8	29.17	9.56	37.65	2284.44
650	25.46	28.77	8.94	37.11	2287.61
700	28.71	27.40	8.16	36.32	2263.00
750	30.18	26.63	8.33	34.2	2292.14
800	32.36	27.30	8.11	31.7	2349.50
850	33.05	26.95	8.82	30.3	2417.53
900	34.00	26.50	8.56	29.00	2406.10
950	34.49	25.89	8.28	29.09	2376.31
1000	35.58	25.25	8.37	28.91	2392.75

Table- A- 2 (c): Effect of temperature on Syn-gas composition for saw dust

(S/B= 0, Equivalence Ratio =0.25, Feed Rate =10kg/hr, Bed material= Silica Sand)

Temp(^oC)	H_2	CO_2	CH_4	CO	NHV Kcal/m^3
500	19.93	21.52	10.18	44.10	2713.63
550	21.29	20.61	9.13	42.15	2600.12
600	25.03	19.85	8.97	39.60	2605.70
650	27.72	19.08	8.52	38.24	2595.15
700	29.98	18.92	8.21	37.62	2608.39
750	31.34	18.75	7.76	35.97	2554.36
800	33.29	18.58	7.60	34.32	2541.20
850	34.80	19.01	7.58	34.16	2573.80
900	35.65	19	7.05	34.11	2548.47
950	37.06	19.00	6.82	34.00	2561.85
1000	39.64	19.14	6.06	33.24	2540.18

Table- A-2 (d): Effect of temperature on Syn-gas composition for wood chips

(S/B= 0, Equivalence Ratio =0.25, Feed Rate =10kg/hr, Bed material= Silica Sand)

Temp (^oC)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	24.42	28.43	8.62	36.39
550	26.24	27.52	8.43	34.42
600	29.37	25.63	7.59	33.12
650	33.18	24.42	8.12	31.81
700	36.63	23.61	7.56	29.78
750	39.84	22.81	7.51	26.42
800	41.16	21.06	8.03	26.73

Table- A-2 (e): Effect of temperature on Syn-gas composition for Sugarcane bagasse(S/B= 0, Equivalence Ratio =0.25, Feed Rate =10kg/hr, Bed material= Silica Sand)

Temp (°C)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	28.02	48.18	12.11	11.04
550	28.17	48.12	12.06	11.02
600	29.34	47.16	11.21	11.01
650	29.29	47.32	11.04	10.81
700	30.16	46.17	10.64	10.72
750	31.24	46.31	10.31	10.44
800	32.42	45.98	9.54	10.16

Table- A-2 (f): Effect of temperature on Syn-gas composition for coconut coir (S/B= 0, Equivalence Ratio =0.25, Feed Rate =10kg/hr, Bed material= Silica Sand)

Temp (^oC)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	26.46	40.81	16.32	15.04
550	27.84	38.92	15.94	13.89
600	28.24	37.62	15.38	12.96
650	28.38	35.87	14.62	11.75
700	29.64	33.86	13.94	11.32
750	30.38	31.24	13.56	10.62
800	31.69	30.63	12.65	9.87

Table- A-3 (a) :Effect of steam to biomass ratio on Syngas composition for Wood chips (Feed rate =10Kg/hr, Temperature = 700°C, Equivalence ratio = 0.25, Bed material= silica sand)

S:B Ratio	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
0.50	31.41	41.34	10.12	16.02
0.75	32.06	41.20	9.81	15.84
1.00	32.37	40.67	9.84	15.24
1.25	33.21	39.24	8.52	14.74

Table- A-3 (b) :Effect of steam to biomass ratio on Syngas composition for Sugarcane bagasse(Feed rate =10Kg/hr, Temp= 700°C, Equivalence ratio = 0.25, Bed material= silica sand)

S:B Ratio	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
0.5	34.34	39.27	11.43	14.23
0.75	35.29	38	11.14	13.68
1	35.02	38	10.67	12.34
1.25	36.28	37	10.24	12.08

Table- A-3 (c) :Effect of steam to biomass ratio on Syngas composition for coconut coir
(Feed rate =10Kg/hr, Temperature = 700°C, Equivalence ratio = 0.25, Bed material= silica sand)

S:B Ratio	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
0.5	32.54	40.28	12.36	13.46
0.75	33.72	39.42	11.84	13.04
1	37.64	37.42	10.68	12.72
1.25	38.89	36.48	10.14	11.72

Table-A-3(d) Effect of steam/biomass ratio on syn-gas composition for rice straw
(Feed rate =10Kg/hr, Temperature = 700°C, Equivalence ratio = 0.25, Bed material= silica sand)

S:B Ratio	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
0.5	29.12	28.91	7.88	33.1
1	30.15	26.33	7.29	30.78
1.5	32.36	27.30	8.11	29.7
2	42.14	23.98	7.12	27.4
2.5	43.54	21.05	6.85	26

Table-A-3(e) Effect of steam/biomass ratio on syn-gas composition for saw dust
(Feed rate =10Kg/hr, Temperature = 700°C, Equivalence ratio = 0.25, Bed material= silica sand)

S:B Ratio	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
0.5	32.12	28.91	7.88	30.1
1	35.15	26.33	7.29	29
1.5	38.29	26.00	7.60	28.00
2	42.14	22.98	7.12	28.4
2.5	43.54	19.05	6.85	28

Table- A-3 (f) Effect of steam/biomass ratio on syn-gas composition for rice husk
(Feed rate =10Kg/hr, Temperature = 700°C, Equivalence ratio = 0.25, Bed material= silica sand)

S:B Ratio	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
0.5	35.12	28.91	7.88	28.1
1	35.15	26.33	7.29	27.78
1.5	38.83	24.50	8.74	27.83
2	42.14	22.98	7.12	25.40
2.5	43.54	19.05	6.85	28.02

Table- A- 4 (a): Effect of Equivalence ratio on Syngas composition for Wood Chips
(Feed rate =10Kg/hr, Temp. = 700°C, steam to biomass ratio = 0.5, Bed material= silica sand)

ER Ratio	H2 (vol %)	CO (vol %)	CH4 (vol %)	CO2 (vol %)
0.2	32.14	44.28	9.62	13.28
0.25	30.24	43.62	8.54	16.24
0.3	27.62	43.27	8.21	19.87
0.35	26.54	42.02	7.69	23.04

Table- A- 4 (b): Effect of Equivalence ratio on Syngas composition for Sugarcane bagasse
(Feed rate =10Kg/hr, Temp. = 700°C, steam to biomass ratio = 0.5, Bed material= silica sand)

ER Ratio	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
0.2	35.24	41.42	11.14	11.21
0.25	34.82	41.29	10.21	14.02
0.3	32.14	40.61	9.84	16.28
0.35	30.21	39.82	9.52	19.68

Table- A- 4 (c): Effect of Equivalence ratio on Syngas composition for coconut coir
(Feed rate =10Kg/hr, Temp. = 700°C, steam to biomass ratio = 0.5, Bed material= silica sand)

ER	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
0.2	32.42	43.62	12.28	11.42
0.25	31.82	42.62	11.12	13.89
0.3	30.52	42.14	10.62	15.38
0.35	30.21	41.72	9.57	17.67

Table- A- 4 (d): Effect of Equivalence ratio on Syngas composition for rice husk
(Feed rate =10Kg/hr, Temp. = 700°C, steam to biomass ratio = 0.5, Bed material= silica sand)

ER	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
0.15	28.71	27.40	8.16	32.32
0.2	30.18	26.63	8.33	30.2
0.25	32.36	27.30	8.11	29.7
0.3	33.05	26.95	8.82	30.3
0.35	34.49	25.89	8.28	32.09

Table- A- 4 (e): Effect of Equivalence ratio on Syngas composition for rice straw
(Feed rate =10Kg/hr, Temp. = 700°C, steam to biomass ratio = 0.5, Bed material= silica sand)

ER	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
0.15	29.98	18.92	8.21	34.62
0.2	31.34	18.75	7.76	33.97
0.25	33.29	18.58	7.60	33.32
0.3	34.80	19.01	7.58	34.16
0.35	37.06	19.00	6.82	35.00

Table- A- 4 (f): Effect of Equivalence ratio on Syngas composition for saw dust
(Feed rate =10Kg/hr, Temp.= 700°C, S/B= 0.5, Bed material= silica sand)

Equivalence Ratio	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
0.15	36.33423	26	7.89518	31.28994
0.2	37.17677	25.6	7.950612	29.04268
0.25	38.8393	25.85	7.74116	26.887
0.3	37.77858	25.33	8.51177	27.14338
0.35	36.05569	22.85	6.1147	33.28274

B. Effect of bed materials (Catalytic effects) on Syngas

Table- B-1 (a) :Temperature effect on Syngas composition for Rice huskwith 1:2 dolomite-sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^{0}C)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	32.15	43.04	12.85	11.57
550	33.16	42.65	12.42	11.59
600	34.12	42.15	12.05	11.63
650	35.1	41.32	11.87	11.67
700	36.25	40.45	11.53	11.69
750	37.05	39.25	11.27	11.73
800	37.95	38.32	11.03	11.76

Table- B-1 (b) :Temperature effect on Syngas composition for Rice huskwith 1:1 dolomite-sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^oC)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	34.12	41.24	13.54	10.72
550	34.87	40.57	13.05	10.74
600	35.48	39.75	12.74	10.77
650	36.02	39.05	12.27	10.82
700	36.64	38.27	11.95	10.86
750	37.57	37.65	11.53	10.92
800	38.25	36.92	11.21	10.98

Table-B-1(c) :Temperature effect on Syngas composition for Rice husk with 1:1 red mud-sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^oC)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	35.54	39.06	16.24	8.86
550	36.24	38.76	15.56	8.96
600	36.92	38.54	14.82	9.12
650	37.87	38.32	13.72	9.27
700	38.42	38.17	12.57	9.35
750	39.87	37.87	11.69	9.49
800	40.25	37.43	11.02	9.58

Table- B-1 (d) :Temperature effect on Syngas composition for Rice husk with 1:2 red mud-sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^oC)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	37.42	37.36	16.24	8.86
550	38.15	36.87	15.38	9.27
600	39.24	36.05	14.92	9.54
650	39.97	35.87	14.12	9.68
700	40.82	35.25	13.78	9.84
750	41.56	34.78	13.19	10.17
800	42.82	33.83	12.82	10.32

Table- B-2 (a) : Temperature effect on Syngas composition for Sugarcane bagasse with 1:2 dolomite -sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^oC)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	31.25	45.02	13.24	10.12
550	32.47	44.31	12.86	10.24
600	33.82	43.62	12.27	10.28
650	34.54	43.03	11.82	10.34
700	35.86	42.06	11.47	10.43
750	36.72	41.24	11.03	10.52
800	37.95	40.12	10.82	10.63

Table- B-2 (b) : Temperature effect on Syngas composition for Sugarcane bagasse with 1:1 dolomite -sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^oC)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	34.27	41.23	14.24	9.82
550	35.12	40.96	13.57	9.89
600	36.07	40.54	12.86	9.97
650	36.92	40.11	12.17	10.07
700	37.87	39.76	11.73	10.13
750	39.12	39.12	11.21	10.21
800	40.27	38.17	10.86	10.26

Table- B-2 (c) : Temperature effect on Syngas composition for Sugarcane bagasse with 1:2red mud -sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^oC)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	37.24	36.34	18.62	7.36
550	37.96	36.98	18.12	7.58
600	38.57	35.42	17.78	7.94
650	39.15	35.02	17.22	8.32
700	39.87	34.57	17.78	8.48
750	40.47	34.25	16.24	8.69
800	40.92	34.03	15.82	8.93

Table- B-2 (d) : Temperature effect on Syngas composition for Sugarcane bagasse with 1:1 red mud -sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^oC)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	38.12	35.54	19.21	6.86
550	38.92	34.78	19.05	6.98
600	39.47	34.35	18.72	7.2
650	39.86	34.03	18.48	7.35
700	40.57	33.44	18.32	7.45
750	41.12	33.18	17.92	7.58
800	41.98	32.51	17.68	7.72

Table- B-3 (a) : Temperature effect on Syngas composition for Coconut coir with 1:2Dolomite -Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^oC)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	30.12	36.27	14.75	18.82
550	30.96	35.52	14.31	18.12
600	31.57	34.89	13.91	17.56
650	32.27	34.12	13.28	17.15
700	33.41	33.43	12.78	16.53
750	34.16	32.82	12.12	16.02
800	35.91	31.95	11.81	15.62

Table- B-3 (b) : Temperature effect on Syngas composition for Coconut coir with 1:1 Dolomite - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^oC)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	32.26	44.18	15.23	8.12
550	32.92	43.78	14.95	7.87
600	33.47	43.21	14.78	7.61
650	33.89	42.83	14.39	7.38
700	34.41	42.57	14.13	7.13
750	34.78	42.18	14.01	6.89

800	35.34	41.96	13.81	6.64

Table- B-3 (c) : Temperature effect on Syngas composition for Coconut coir with 1: 2 Red mud - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^oC)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	34.78	42.27	12.62	9.92
550	35.52	41.72	12.28	9.87
600	36.27	41.18	12.03	9.78
650	36.98	40.78	11.81	9.62
700	37.44	40.34	11.53	9.41
750	37.89	39.57	11.27	9.34
800	38.38	38.76	10.93	9.22

Table- B-3 (d) :Temperature effect on Syngas composition for Coconut coir with 1: 1 Red mud - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temperature (^oC)	H_2(vol%)	CO (vol%)	CH_4 (vol%)	CO_2(vol%)
500	37.42	39.71	14.12	8.24
550	37.97	39.31	13.78	8.07
600	38.52	39.05	13.53	7.91
650	38.89	38.68	13.34	7.87
700	39.47	38.24	13.12	7.81
750	39.72	37.85	12.87	7.73
800	40.28	37.52	12.67	7.62

Table- B-4 (a) : Temperature effect on Syngas composition for wood chips with 1:2 Dolomite -Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron

Temp (^oC)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	29.35	26.13	9.21	35.31
550	30.14	25.52	13.22	31.12
600	31.37	24.23	14.28	30.12
650	34.15	23.45	12.72	29.68
700	38.26	21.54	13.02	27.18
750	40.43	20.14	14.66	24.77
800	42.56	19.06	13.06	25.32

Table- B-4 (b) : Temperature effect on Syngas composition for wood chips with 1:1 Dolomite - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temp (^oC)	H2 (vol %)	CO (vol %)	CH4 (vol %)	CO2 (vol %)
500	28.52	26.43	10.86	34.19
550	29.24	25.32	13.21	32.23
600	31.87	23.36	13.55	31.22
650	32.28	22.24	16.64	28.84
700	39.23	20.17	12.78	27.82
750	39.88	20.81	13.59	25.72
800	42.32	19.56	13.99	24.13

Table- B-4 (c) : Temperature effect on Syngas composition for wood chips with 1:2redmud - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temp (^oC)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	33.45	26.43	5.73	34.39
550	34.45	24.52	8.61	32.42
600	34.78	24.13	9.33	31.76
650	35.88	23.52	11.47	29.13
700	37.34	21.61	13.57	27.48
750	40.78	20.81	12.99	25.42
800	42.32	19.56	13.39	24.73

Table- B-4 (d) : Temperature effect on Syngas composition for wood chips with 1:1 redmud - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temp (^oC)	H_2 (vol %)	CO (vol %)	CH_4 (vol	CO_2 (vol %)
500	32.46	25.43	8.88	33.23
550	33.12	24.52	10.24	32.12
600	33.78	23.63	11.36	31.23
650	35.38	22.42	13.02	29.18
700	38.68	22.12	11.93	27.27
750	41.24	20.45	12.93	25.38
800	43.25	20.06	12.36	24.33

Table- B-5 (a) : Temperature effect on Syngas composition for rice strawwith 1:1 Dolomite - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temp (°C)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	20.25	30.32	11.23	38.2
550	22.47	31.8	8.23	37.5
600	24.48	30.37	8.62	36.53
650	27.48	30.87	5.84	35.81
700	30.41	31.45	3.32	34.82
750	31.45	28.33	8.02	32.2
800	33.63	29.13	6.57	30.67

Table- B-5 (b) : Temperature effect on Syngas composition for rice strawwith 1:2 Dolomite - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temp (°C)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	19.53	31.3	10.77	38.4
550	21.75	30.8	9.95	37.5
600	23.8	29.17	10.38	36.65
650	26.44	28.77	8.68	36.11
700	29.79	27.4	7.49	35.32
750	31.15	26.63	9.02	33.2
800	32.9	27.3	9.1	30.7

Table- B-5 (c) : Temperature effect on Syngas composition for rice strawwith 1:2redmud - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temp (°C)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	19.75	29.65	12.37	38.23
550	21.57	28.89	12.98	36.56
600	24.15	27.57	13.63	34.65
650	27	26.77	10.56	35.67
700	31.73	25.57	9.2	35.52
750	32.18	24.13	10.41	34.28
800	33.1	25.3	11.33	30.27

Table- B-5 (d) : Temperature effect on Syngas composition for rice strawwith 1:1redmud - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temp (°C)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	20.85	30.13	10.02	39
550	22.55	29.8	9.65	38
600	24.8	28.12	9.43	37.65
650	27.26	27.67	7.96	37.11
700	32.45	25.4	5.83	36.32
750	33.34	24.53	7.93	34.2
800	34.36	26.53	9.41	29.7

Table- B-6 (a) : Temperature effect on Syngas composition for sawdust with 1:1 Dolomite - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temp (°C)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	20.13	20.15	17.62	42.1
550	22.32	19.61	17.92	40.15
600	26.13	17.85	18.52	37.5
650	28.82	20.23	14.61	36.34
700	30.56	17.32	17	35.12
750	32.78	17.65	15.7	33.87
800	34.49	17.8	14.39	33.32

Table- B-6 (b) : Temperature effect on Syngas composition for sawdust with 1:2 Dolomite - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temp (°C)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	20	21.54	15.36	43.1
550	21.78	19.5	17.57	41.15
600	25.83	17.8	20.17	36.2
650	28.32	17	18.46	36.22
700	31.98	18.85	13.75	35.42
750	32.14	18.51	15.88	33.47
800	32.29	18.45	16.96	32.3

Table- B-6 (c) : Temperature effect on Syngas composition for sawdust with 1:2redmud - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temp (°C)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	21.22	21.85	14.77	42.16
550	21.46	21.61	16.58	40.35
600	26.03	20	15.97	38
650	28.72	19.85	13.98	37.45
700	32.12	19.12	12.5	36.26
750	32.34	19.35	13.34	34.97
800	34.29	19.58	12.81	33.32

Table- B-6 (d) : Temperature effect on Syngas composition for sawdust with 1:1redmud - Sand bed material (ER= 0.25, S/B=0.5, Dolomite particle size : 400-500micron)

Temp (°C)	H_2 (vol %)	CO (vol %)	CH_4 (vol %)	CO_2 (vol %)
500	21.93	21.52	14.45	42.1
550	22.69	20.61	16.55	40.15
600	26.86	19.85	14.69	38.6
650	29.12	20.08	13.56	37.24
700	33	18.92	11.46	36.62
750	33.34	18.75	12.94	34.97
800	35.29	18.58	11.31	34.82

Appendix-II

(A) Energy Balance and Mass Balance Calculations for Coconut Coir

Data:

Biomass Fuel - 1 kg/hr (0.947 kg/hr dry)

Steam – 0.5 kg/kg fuel

Moisture is 5.3%

Composition of fuel on %dry basis

H_2 =4.96, C = 43.76, N_2 =0.4, O_2 = 47.2, S = 0.22

Product gas:

H_2 =10.14%, CH_4 = 4.05%, CO = 9.8%, CO_2 = 3.16%, N_2 = 49% and O_2=19%

HHV=19 MJ/kg

From experiment air flow rate is noted to be = 0.00767m^3/hr

Specific humidity of supply air be 0.01kg moisture per kg dry air

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Basis: 1 kg of fuel with 0.053kg moisture, 0.947kg dry fuel

Air: Flow rate = 0.00767m^3/hr (corresponding to Equivalence Ratio=0.2) for Feed rate of 10kg/hr

Thus air flow rate for 1kg/hr of feed is calculated as = 2.76kg/kg of fuel

Steam: 0.5 kg/kg of fuel (since steam rate is 5 kg/hr for 10kg/hr of feed)

Air contains 76.9% N_2 by mass

So N_2 from air

= 0.769* 2.76 = 2.12244 kg N_2/kg fuel

Total N_2 supplied by air and fuel which carry 0.40% N_2

= 2.12244+ 0.004 = 2.12644 kg N_2 /kg fuel

= 0.075944285kg mol/ kg fuel

We know product gas contains 49% N_2 . Hence the amount of product gas produced

= 0.075944285/ 0.49 = 0.154988338 kg mol gas / kg fuel

Product Gas=0.155 kg mol / kg fuel

Oxygen inflow to Gasifier :

Oxygen supplied to gasifier (with air) is

= 0.2315 * 2.76 = 0.63894 kg/kg fuel

Steam supplied = 0.5 kg /kg fuel

So oxygen associated with steam supply

= 16/18 *0.5 = 0.4444 kg /kg fuel

Moisture in fuel = 5.3%

Oxygen with moisture in fuel

= 0.053 * 16/18 = 0.0471111 kg/kg fuel

Oxygen with moisture in air supply

= 0.01 * 2.76 * 16/18 = 0.024533 kg/kg fuel

Oxygen in fuel =47.2% =0.472 kg/kg fuel

Total oxygen flow to gasifier with air steam, moisture in fuel and air

= 0.63894 + 0.44444 + 0.0471111 + 0.024333 + 0.472

= 1.6270283 kg/kg fuel

≈ **1.627 kg/kg fuel**

H_2 Balance:

Total H_2 in flow to the gasifier with fuel, steam, moisture in fuel and moisture in air

= 0.058 * 0.947 + 0.5 * 2/18 + 0.053 * 2/18 + 0.01* 2.76*2/18

= 0.119437111 kg/kg fuel

H_2 associated with H_2 & CH_4 in dry product gas

= (0.1014 + 0.0405 * 2) * 0.155

= 0.028272 kg mol/kg fuel

=0.056 kg H_2 /kg fuel

Assuming all S converted to H_2S and removed by the gas cleaning system, hydrogen associated with H_2S in the raw product gas

=0.0022 * 2/32 = 1.375 * 10^{-4} kg/kg fuel

Total hydrogen in the uncleaned dry product gas including that in H_2S

=0.056544 + 1.375 * 10^{-4}= 0.0566815 kg/kg fuel

= 0.057 kg H_2 /kg fuel

To find the moisture in the product gas we deduct the hydrogen in dry gas from the total hydrogen in flow obtained earlier using the hydrogen balance.

= Hydrogen in flow – Hydrogen in dry gas

= 0.119437111 – 0.056682

= 0.062755111 ≈ 0.063 kg/kg fuel

Steam associated with this hydrogen in the gas

=0.062755111* 18/2 = 0.564795999 = **0.565 kg /kg fuel**

<u>Oxygen Balance:</u>

Oxygen associated with CO, CO_2 in dry gas which have half a mol and 1 mol of oxygen respectively.

= (0.5 * 0.098 + 0.0316+0.19) * 0.155

= 0.2706*0.155

=0.041943 kg mol /kg fuel

= 0.041943 * 32 = 1.342176 kg /kg fuel

Oxygen associated with the steam in gas

= 0.5648 * 16/18 =0.502044444 kg /kg fuel

Total Oxygen in gas = 1.342176 + 0.502044444 = 1.844220444 =**1.844 kg/kg fuel**

Here we note that this is slightly more than the oxygen in flow of **1.627 kg /kg fuel** calculated earlier. This must be due to measurement errors in the given data on fuel gas composition.

Carbon Balance:

Total carbon associated with CO, CO_2 and CH_4 in dry gas whose production rate has been computed earlier as 0.152 kg-mol / kg fuel is

= (0.098 + 0.0316 + 0.0405) * 0.155

= 0.1701 * 0.155 = 0.0263655 kg mol/ kg

= 0.0263655 * 12 = **0.316386= 0.316 kg/kg fuel**

Carbon input is found from the composition of the fuel

= **0.384kg /kg fuel**

Carbon conversion efficiency

= (0.3164/0.3845) * 100

= 0.822887*100 =82.3%

Energy Analysis

Components	**Heats of Combustion, MJ/kg mol**
CO	**282.99**
H_2	**285.84**
CH_4	**890.36**

Energy output with CO

=0.098*0.155 (kg mol CO/kg fuel)*282.99= 4.29862 MJ/kg fuel

Energy output with CH_4

=0.0405*0.155(kg mol CH_4/kg fuel)*890.36= 5.58 9235MJ/kg fuel

Energy output with H_2

=0.1014*0.155 (kg molH_2/kg fuel)*285.84= 4.4925473 MJ/kg

Total Energy output

=4.29862 +5.58 9235+4.4925473 **=14.380 MJ/kg fuel**

Total energy input is 19 MJ/kg fuel

Cold gas efficiency of the gasifier

=(14.3804 /19)*100=75.686%

Proposed Energy Audit report

Energy input: As already discussed in total energy input is 8.26kWh

Energy output

As per material balance calculation it is found that 1kg of biomass sample produces 0.155kmol gas which occupies =0.155*22.4=3.472m^3at STP. This implies that vol. of gas produced at 800^0C will be =3.472*(1073/273) =13.65 m^3, this is under ideal conditions. But the gas analyser reads product gas as 10 m^3/kg of biomass samples. This difference in yield might be due to the fact that the gasifier is not 100% efficient for which all the biomass are not gasified completely and some losses of product gas might be there.

Energy output was calculated using the heating value of the flue gases.

Flow rate of the flue gases from the gasifier unit is1.0 m^3/ hr for 1kg of fuel. As feed rate is 10kg/hr, flow rate of gas can be said as 10 m^3/ hr.

Composition of the flue gas for air gasification is as follows

	O_2	CO	CO_2	CH_4	H_2	N_2
Mole or, Vol%	**19.0**	**9.8**	**3.16**	**4.05**	**10.14**	**49.0**
Flow rate(m^3/hr)	**1.90**	**0.98**	**0.316**	**0.405**	**1.014**	**4.90**

Temperature of the flue gases coming out is 800^0C

The heat energy processed by the flue gases was sapped out before calculating the net heating value of the flue gases.

The heat energy extracted from the flue gases:(density of flue gas =0.345kg/$m^3$and mean specific heat at 900^0C is 3.735KJ/kg.^{0}K)

Heat Energy:$\sum mc_p$*(800-35) = (10*0.345)*3.735*765=12.886 *765= 9857.59875 kJ/kg

= 9.858 MJ/hr=2.738243781 kWhr/hr=**2.74 kW(approx.)**

Energy expected to be obtained from the flue gases will be

Compound	Heating Value, (KJ/m^3)
CO	4074
CH_4	11605
H_2	3450

Output energy expected to be obtained from the different components of the flue gas:

(0.98 *4074) + (0.405*11605) + (1.014*3450) = 12190.845 kJ/hr = 12.191 MJ/hr

= 3.386372924 kWhr/hr= 3.3864 kW (**approx.)**

Total Energy output = 2.74 + 3.3864= 6.1264kWh per kg of fuel

Net energy output= 6.1264*10 –8.26 = 53.004 kWh

10 kg of feed (dry basis) produces53.004 kWh energy equivalent of synthesis gas and 10 $Nm^3$of gas per hour.

Typical Operation Yields and Efficiencies

The operation results give the following yields and efficiencies:

1 kg of feed (dry basis) yielded 0.413kWhr equivalent of synthesis Gas.

1 kg of feed yielded 1.0 Nm^3 of gas with 10.14% H_2, 4.05% CH_4, 9.8% CO, 3.16% CO_2 and 19% $O_2$and 49% N_2.

Overall gasification efficiency was 50% (considering LHV of feed to LHV of GasOutput).

(B) Energy Balance and Mass Balance Calculations for wood chips

Data:

Biomass Fuel - 1 kg/hr (0.92 kg/hr dry)

Steam – 0.5 kg/kg fuel

Moisture is 8%

Composition of fuel on %dry basis

H_2 =5.7, C = 46.23, N_2 =0.22, O_2 = 45.2, S = 0.12

Product gas:

H_2 =14.17%, CH_4 = 2.57%, CO = 8.74%, CO_2 = 6.55%, N_2 = 50% and O_2=18%

HHV=15.6MJ/kg

From experiment air flow rate is noted to be = 0.00767m^3/hr

Specific humidity of supply air be 0.01kg moisture per kg dry air

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Basis: 1 kg of fuel with0.8 kg moisture, 0.92kg dry fuel

Air: Flow rate = 0.00767m^3/hr (corresponding to Equivalence Ratio=0.2) for Feed rate of 10kg/hr

Thus air flow rate for 1kg/hr of feed is calculated as = 2.76kg/kg of fuel

Steam: 0.5 kg/kg of fuel (since steam rate is 5 kg/hr for 10kg/hr of feed)

Air contains 76.9% N_2 by mass

So N_2 from air

= 0.769* 2.76 = 2.12244 kg N_2/kg fuel

Total N_2 supplied by air and fuel which carry 0.22% N_2

= 2.12244+ 0.0022= 2.1266 kg N_2 /kg fuel

= 0.07595kg mol/ kg fuel

We know product gas contains 50% N_2. Hence the amount of product gas produced

= 0.07595/ 0.50 = 0.1519 kg mol gas / kg fuel

Product Gas=0.152 kg mol / kg fuel

Oxygen inflow to Gasifier :

Oxygen supplied to gasifier (with air) is

= 0.2315 * 2.76 = 0.6389 kg/kg fuel

Steam supplied = 0.5 kg /kg fuel

So oxygen associated with steam supply

= 16/18 *0.5 = 0.4444 kg /kg fuel

Moisture in fuel = 8%

Oxygen with moisture in fuel

= 0.08 * 16/18 = 0.07111 kg/kg fuel

Oxygen with moisture in air supply

= 0.01 * 2.76 * 16/18 = 0.024533 kg/kg fuel

Oxygen in fuel =45.2% =0.452g/kg fuel

Total oxygen flow to gasifier with air steam, moisture in fuel and air

= 0.63894 + 0.44444 + 0.07111+ 0.024533 + 0.452

= 1.63103 kg/kg fuel

≈ **1.631 kg/kg fuel**

H_2 Balance:

Total H_2 in flow to the gasifier with fuel, steam, moisture in fuel and moisture in air

= 0.057 * 0.92 + 0.5 * 2/18 + 0.08 * 2/18 + 0.01* 2.76*2/18

= 0.1199511 kg/kg fuel

H_2 associated with H_2 & CH_4 in dry product gas

= (0.1417 + 0.0257 * 2) * 0.1519

= 0.0293 kg mol/kg fuel

=0.059 kg H_2 /kg fuel

Assuming all S converted to H_2S and removed by the gas cleaning system, hydrogen associated with H_2S in the raw product gas

=0.0012 * 2/32 = 7.5 * 10^{-3} kg/kg fuel

Total hydrogen in the un-cleaned dry product gas including that in H_2S

=0.0586 + 7.5 * 10^{-3}= 0.058675 kg/kg fuel

= 0.057kg H_2 /kg fuel

To find the moisture in the product gas we deduct the hydrogen in dry gas from the total hydrogen in flow obtained earlier using the hydrogen balance.

= Hydrogen in flow – Hydrogen in dry gas

= 0.1199511 – 0.058675

= 0.061276 kg/kg fuel

Steam associated with this hydrogen in the gas

=0.061276 * 18/2 = **0.552 kg /kg fuel**

Oxygen Balance:

Oxygen associated with CO, CO_2 in dry gas which have half a mol and 1 mol of oxygen respectively.

= (0.5 * 0.0874 + 0.0655+0.18) * 0.1519

= 0.016587 kg mol /kg fuel

= 0.016587 * 32 = 0.531 kg /kg fuel

Oxygen associated with the steam in gas

= 0.592524 * 16/18 =0.526688 kg /kg fuel

Oxygen in gas= 0.18*0.1519*32=0.874944

Total Oxygen in gas = 0.531+0.526688+0.874944 = 1.932632 = **1.933 kg/kg fuel**

Here we note that this is slightly more than the oxygen in flow of **1.6310 kg /kg fuel** calculated earlier. This must be due to measurement errors in the given data on fuel gas composition.

Carbon Balance:

Total carbon associated with CH_4, CO_2andCO in dry gas whose production rate has been computed earlier as 0.1519 kg-mol / kg fuel is

= (0.0257+ 0.0655 + 0.0874) * 0.1519

=0.02713kg mol/ kg

= 0.02713 * 12 = **0.325 kg/kg fuel**

Carbon input is found from the composition of the fuel

= 0.4623**kg /kg fuel**

Carbon conversion efficiency

= (0.3255521/0.4623) * 100

= 0.7042=70.42%

Energy Analysis

Components	**Heats of Combustion, MJ/kg mol**
CO	**282.99**
H_2	**285.84**
CH_4	**890.36**

Energy output with CO

=0.0874*0.1519(kg mol CO/kg fuel)*282.99=3.75699 MJ/kg fuel

Energy output with CH_4

=0.0257*0.1519 (kg molCH_4/kg fuel)*890.36=3.4758MJ/kg fuel

Energy output with H_2

=0.1417*0.1519 (kg molH_2/kg fuel)*285.84=6.1525MJ/kg

Total Energy output

=3.75699+3.4758+6.1525**=13.385MJ/kg fuel**

Total energy input is 15.6 MJ/kgg fuel

Cold gas efficiency of the gasifier

=(13.39/15.6)*100=85.8%

Proposed Energy Audit report

Energy input =8.26kWh (already calculated)

Energy output

Energy output was calculated using the heating value of the flue gases.

As per material balance calculation it is found that 1kg of biomass sample produces 0.152kmol gas which occupies =0.152*22.4=3.4048m^3at STP. This implies that vol. of gas produced at 800^0C will be =3.4048*(1073/273) =13.382 m^3, this is under ideal conditions. But the gas analyser reads product gas as 10 m^3/kg of biomass samples. This difference in yield might be due to the fact that the gasifier is not 100% efficient for which all the biomass are not gasified completely and some losses of product gas might be there.

Composition of the flue gas for air gasification is as follows

	O_2	**CO**	**CO_2**	**CH_4**	**H_2**	**N_2**
Mole or, Vol%	**18**	**8.74**	**6.55**	**2.57**	**14.17**	**50**
Flow rate(m^3/hr)	**1.8**	**0.874**	**0.655**	**0.257**	**1.417**	**0.5**

Temperature of the flue gases coming out is 800^0C

The heat energy processed by the flue gases was sapped out before calculating the net heating value of the flue gases.

The heat energy extracted from the flue gases:(density of flue gas =0.345kg/$m^3$and mean specific heat at 900^0C is 3.735KJ/kg.^{0}K)

Heat Energy:$\sum mc_p$*(800-35) = (10*0.345)*3.735*765=12.886 *765= 9857.59875 kJ/kg

= 9.858 MJ/hr=2.738243781 kWhr/hr=**2.74 kW(approx.)**

Energy expected to be obtained from the flue gases will be

Compound	Heating Value, (KJ/m^3)
CO	4074
CH_4	11605
H_2	3450

Output energy expected to be obtained from the different components of the flue gas:

(0.874 *4074) + (0.257*11605) + (1.417*3450) = 11431.811kJ/hr = 11.4318 MJ/hr

= 3.17553 kWhr/hr= 3.176 kW (**approx.)**

Total Energy output = 2.74+ 3.176 = 5.916kWh per kg of fuel

Net energy output= 5.916*10 – 8.26 = 50.9 kWh=51kWh

10 kg of feed (dry basis) produces51 kWh energyequivalent of synthesis gas and 10 $Nm^3$of gas per hour.

Typical Operation Yields and Efficiencies

The operation results give the following yields and efficiencies:

1 kg of feed (dry basis) yielded 0.425kWhr equivalent of synthesis Gas.

1 kg of feed yielded 1.0 Nm^3 of gas with 14.17% H_2, 2.57% CH_4, 8.74% CO, 6.55% CO_2 and 18% $O_2$and 50% N_2.

Overall gasification efficiency was 50% (considering LHV of feed to LHV of GasOutput).

(C) Energy Balance and Mass Balance Calculations for Rice Husk

Data:

Biomass Fuel - 1 kg/hr (0.9266 kg/hr dry)

Steam – 0.5 kg/kg fuel

Moisture is 7.34%

Composition of fuel on %dry basis

H_2 =4.96, C = 38.45, N_2 =0.82, O_2 = 55.59, S = 0.18

Product gas:

H_2 =17.56%, CH_4 = 2.9%, CO = 9.2%, CO_2 = 6.8%, N_2 = 48.2% and O_2=15.34%

HHV=16.78MJ/kg

From experiment air flow rate is noted to be = 0.00767m^3/hr

Specific humidity of supply air be 0.01kg moisture per kg dry air

---

Basis: 1 kg of fuel with 0.734kg moisture, 0.9266kg dry fuel

Air: Flow rate = 0.00767m^3/hr (corresponding to Equivalence Ratio=0.2) for Feed rate of 10kg/hr

Thus air flow rate for 1kg/hr of feed is calculated as = 2.76kg/kg of fuel

Steam: 0.5 kg/kg of fuel (since steam rate is 5 kg/hr for 10kg/hr of feed)

Air contains 76.9% N_2 by mass

So N_2 from air

= 0.769* 2.76 = 2.12244 kg N_2/kg fuel

Total N_2 supplied by air and fuel which carry 0.82% N_2

= 2.12244+ 0.0082 = 2.13064 kg N_2 /kg fuel

= 0.0761 kg mol/ kg fuel

We know product gas contains 48.2% N_2 . Hence the amount of product gas produced

= 0.0761/ 0.482 = 0.15788 kg mol gas / kg fuel

Product Gas=0.158kg mol / kg fuel

Oxygen inflow to Gasifier :

Oxygen supplied to gasifier (with air) is

= 0.2315 * 2.76 = 0.6389 kg/kg fuel

Steam supplied = 0.5 kg /kg fuel

So oxygen associated with steam supply

= 16/18 *0.5 = 0.4444 kg /kg fuel

Moisture in fuel = 7.34%

Oxygen with moisture in fuel

= 0.0734 * 16/18 = 0.065244 kg/kg fuel

Oxygen with moisture in air supply

= 0.01 * 2.76 * 16/18 = 0.024533 kg/kg fuel

Oxygen in fuel =36.5% =.365kg/kg fuel

Total oxygen flow to gasifier with air steam, moisture in fuel and air

= 0.63894 + 0.44444 + 0.065244 + 0.024333 + 0.365

= 1.53796 kg/kg fuel

≈ **1.538 kg/kg fuel**

H_2 Balance:

Total H_2 in flow to the gasifier with fuel, steam, moisture in fuel and moisture in air

= 0.0496 * 0.9266 + 0.5 * 2/18 + 0.0734 * 2/18 + 0.01* 2.76*2/18

= 0.1127371377 kg/kg fuel

H_2 associated with H_2 & CH_4 in dry product gas

= (0.1756 + 0.029 * 2) * 0.15788

= 0.03688 kg mol/kg fuel

=0.073 kg H_2 /kg fuel

Assuming all S converted to H_2S and removed by the gas cleaning system, hydrogen associated with H_2S in the raw product gas

=0.0018 * 2/32 = 1.125 * 10^{-4} kg/kg fuel

Total hydrogen in the uncleaned dry product gas including that in H_2S

=0.073761536 + 1.125 * 10^{-4}= 0.07706034 kg/kg fuel

= 0.074 kg H_2 /kg fuel

To find the moisture in the product gas we deduct the hydrogen in dry gas from the total hydrogen in flow obtained earlier using the hydrogen balance.

= Hydrogen in flow – Hydrogen in dry gas

= 0.112737137 – 0.073874036

= 0.038863101 ≈ 0.039 kg/kg fuel

Steam associated with this hydrogen in the gas

=0.038863101* 18/2 = 0.34976792 = **0.350kg /kg fuel**

<u>Oxygen Balance:</u>

Oxygen associated with CO, CO_2 in dry gas which have half a mol and 1 mol of oxygen respectively.

= (0.5 * 0.092 + 0.068+0.1534) * 0.15788

=0.2674*0.15788

= 0.04222 kg mol /kg fuel

= 0.04222 * 32 = 1.35104 kg /kg fuel

Oxygen associated with the steam in gas

= 0.382534 * 16/18 =0.34 kg /kg fuel

Total Oxygen in gas = 1.35104 + 0.34 = =1.69104 = **1.691 kg/kg fuel**

Here we note that this is slightly more than the oxygen in flow of **1.538 kg /kg fuel** calculated earlier. This must be due to measurement errors in the given data on fuel gas composition.

Carbon Balance:

Total carbon associated with CO, CO_2 and CH_4 in dry gas whose production rate has been computed earlier as 0.152 kg-mol / kg fuel is

= (0.029 + 0.068 + 0.092) * 0.15788

= 0.189 * 0.15788 = 0.02984 kg mol/ kg

= 0.02984 * 12 = **0.358 kg/kg fuel**

Carbon input is found from the composition of the fuel

= 0.3845**kg /kg fuel**

Carbon conversion efficiency

= (0.358072/0.3845) * 100

= 0.9312662=93.13%

Energy Analysis

Components	**Heats of Combustion, MJ/kg mol**
CO	**282.99**
H_2	**285.84**
CH_4	**890.36**

Energy output with CO

=0.092*0.15788 (kg mol CO/kg fuel)*282.99=1.2957MJ/kg fuel

Energy output with CH_4

=0.029*0.15788(kg mol CH_4/kg fuel)*890.36=4.07653MJ/kg fuel

Energy output with H_2

=0.1756*0.15788 (kg molH_2/kg fuel)*285.84=7.92455MJ/kg

Total Energy output

=1.2957+4.07653+7.92455=13.297 MJ/kg fuel

Total energy input is 16.2 MJ/kg fuel

Cold gas efficiency of the gasifier

=(13.297/16.2)*100=82.08%

Proposed Energy Audit report

Energy input :8.26kWh (already calculated)

Energy output

As per material balance calculation it is found that 1kg of biomass sample produces 0.158kmol gas which occupies =0.158*22.4=3.54m^3at STP. This implies that vol. of gas produced at 800^{0}C will be =3.54*(1073/273) =13.914 m^3, this is under ideal conditions. But the gas analyser reads product gas as 10 m^3/kg of biomass samples. This difference in yield might be due to the fact that the gasifier is not 100% efficient for which all the biomass are not gasified completely and some losses of product gas might be there.

Energy output was calculated using the heating value of the flue gases.

Flow rate of the flue gases from the gasifier unit 1.1 m^3/ hr for 1kg of fuel as feed rate is 10kg/hr, flow rate of gas can be said as 11 m^3/ hr.

Composition of the flue gas for air gasification is as follows

	O_2	**CO**	**CO_2**	**CH_4**	**H_2**	**N_2**
Mole or, Vol%	**15.34**	**9.2**	**6.8**	**2.9**	**17.56**	**48.2**
Flow rate(m^3/hr)	**1.6874**	**1.012**	**0.748**	**0.319**	**1.9316**	**5.302**

Temperature of the flue gases coming out is 800^{0}C

The heat energy processed by the flue gases was sapped out before calculating the net heating value of the flue gases.

The heat energy extracted from the flue gases:(density of flue gas =0.345kg/$m^3$and mean specific heat at 900^{0}C is 3.735KJ/kg.^{0}K)

Heat Energy:$\sum mc_p*(800-35)$ = (11*0.345)*3.735*765=14.174*765= 10843.3586 kJ/kg

= 10.843 MJ/hr=3.012kWhr/hr=3.012KW**(3.00 kW approx.)**

Energy expected to be obtained from the flue gases will be

Compound	Heating Value, (KJ/m^3)
CO	4074
CH_4	11605
H_2	3450

Output energy expected to be obtained from the different components of the flue gas:

(1.012 *4074) + (0.319*11605) + (1.9316*3450) = 14488.903 KJ/hr = 14.4889 MJ/hr

= 4.02473KWhr/hr= 4.025 kW (**approx.)**

Total Energy output = 3.012 + 4.025= 7.037 kWh per kg of fuel

Net energy output=7.037*10 – 8.26 = 62.11kWh

10 kg of feed (dry basis) produces 62.11kWhr energy equivalent of synthesis gas and 11 $Nm^3$of gas per hour.

Typical Operation Yields and Efficiencies

The operation results give the following yields and efficiencies:

1 kg of feed (dry basis) yielded 0.5kWhr equivalent of synthesis Gas.

1 kg of feed yielded 1.1 Nm^3 of gas with 17.56% H_2, 2.9% CH_4, 9.2% CO, 6.8% CO_2 and

15.34% $O_2$and 48.2% N_2.

Overall gasification efficiency was 50% (considering LHV of feed to LHV of GasOutput).

(D)Energy Balance and Mass Balance Calculations for Rice Straw

Data:

Biomass Fuel - 1 kg/hr (0.9062 kg/hr dry)

Steam – 0.5 kg/kg fuel

Moisture is 9.38%

Composition of fuel on %dry basis

H_2 =4.55, C = 38.6, N_2 =0.47, O_2 =56.17, S = 0.21, H_2O = 9.38

Product gas:

H_2 =14.56%, CH_4 = 2.52%, CO = 9.25%, CO_2 = 8.0%, N_2 = 49.24% and O_2=13.83%

HHV=16.78 MJ/kg

From experiment air flow rate is noted to be = 0.00767m^3/hr

Specific humidity of supply air be 0.01kg moisture per kg dry air

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Basis: 1 kg of fuel with 0.0938kg moisture, 0.9062kg dry fuel

Air: Flow rate = 0.00767m^3/hr (corresponding to Equivalence Ratio=0.2) for Feed rate of 10kg/hr

Thus air flow rate for 1kg/hr of feed is calculated as = 2.76kg/kg of fuel

Steam: 0.5 kg/kg of fuel (since steam rate is 5 kg/hr for 10kg/hr of feed)

Air contains 76.9% N_2 by mass

So N_2 from air

= 0.769* 2.76 = 2.12244 kg N_2/kg fuel

Total N_2 supplied by air and fuel which carry 0.47% N_2

= 2.12244+ 0.0047 = 2.12714 kg N_2 /kg fuel

= 0.076 kg mol/ kg fuel

We know product gas contains 49.24% N_2. Hence the amount of product gas produced

= 0.076/ 0.4924 = 0.1544 kg mol gas / kg fuel

Product Gas=0.154 kg mol / kg fuel

Oxygen inflow to Gasifier :

Oxygen supplied to gasifier (with air) is

= 0.2315 * 2.76 = 0.6389 kg/kg fuel

Steam supplied = 0.5 kg /kg fuel

So oxygen associated with steam supply

= 16/18 *0.5 = 0.4444 kg /kg fuel

Moisture in fuel = 9.38%

Oxygen with moisture in fuel

= 0.0938 * 16/18 = 0.083378 kg/kg fuel

Oxygen with moisture in air supply

= 0.01 * 2.76 * 16/18 = 0.024533 kg/kg fuel

Oxygen in fuel =37.2% = 0.372 kg/kg fuel

Total oxygen flow to gasifier with air steam, moisture in fuel and air

= 0.63894 + 0.44444 + 0.083378 + 0.024333 + 0.372

= 1.5631 kg/kg fuel

≈ **1.563 kg/kg fuel**

H_2 Balance:

Total H_2 in flow to the gasifier with fuel, steam, moisture in fuel and moisture in air

= 0.0455*0.9062 + 0.01 *2.76* 2/18 + 0.5 * 2/18 + 0.0938*2/18

= 0.110276544 kg/kg fuel

H_2 associated with H_2 & CH_4 in dry product gas

= (0.1456 + 0.0252 * 2) * 0.1544

= 0.0302624 kg mol/kg fuel

=0.061 kg H_2 /kg fuel

Assuming all S converted to H_2S and removed by the gas cleaning system, hydrogen associated with H_2S in the raw product gas

=0.12/100 *2/32 = 7.5* 10^{-5} kg/kg fuel

Total hydrogen in the uncleaned dry product gas including that in H_2S

=0.0605248 + 7.5 * 10^{-5}= 0.0605998 kg/kg fuel

= 0.0606 kg H_2 /kg fuel

To find the moisture in the product gas we deduct the hydrogen in dry gas from the total hydrogen in flow obtained earlier using the hydrogen balance.

= Hydrogen in flow – Hydrogen in dry gas

=0.110276544 - 0.0605998

=0.049676744 kg/kg fuel

Steam associated with this hydrogen in the gas

=0.049676644 * 18/2 = 0.4471 = **0.447 kg /kg fuel**

Oxygen Balance:

Oxygen associated with CO, CO_2 in dry gas which have half a mol and 1 mol of oxygen respectively.

= (0.5 * 0.092 + 0.08) * 0.1544

= 0.0194544 kg mol /kg fuel

= 0.0194544 * 32 = 0.623776 kg /kg fuel

Oxygen associated with the steam in gas

= 0.4853898 * 16/18 =0.4314576 kg /kg fuel

Oxygen already available in the gas = 0.1383*0.1544*32 = 0.68331264

Total Oxygen in gas = 0.623776 + 0.4314576+ 0.68331264

= 1.73855 = **1.739 kg/kg fuel**

Here we note that this is slightly more than the oxygen in flow of **1.563 kg /kg fuel** calculated earlier. This must be due to measurement errors in the given data on fuel gas composition.

Carbon Balance:

Total carbon associated with CO, CO_2 and CH_4 in dry gas whose production rate has been computed earlier as 0.152 kg-mol / kg fuel is

= (0.0925+0.08+0.0252) * 0.1544

= 0.030525 kg mol/ kg

= 0.030525 * 12 = **0.366 kg/kg fuel**

Carbon input is found from the composition of the fuel

= 0.386**kg /kg fuel**

Carbon conversion efficiency

= (0.3663/0.386) * 100

= 0.9488 =95.0 %

Energy Analysis

Components	**Heats of Combustion, MJ/kg mol**
CO	**282.99**
H_2	**285.84**
CH_4	**890.36**

Energy output with CO

=0.092*0.1544(kg mol CO/kg fuel)*282.99=4.042MJ/kg fuel

Energy output with CH_4

=0.02522*0.1544(kg molCH_4/kg fuel)*890.36=3.467MJ/kg fuel

Energy output with H_2

=0.1456*0.1544(kg molH_2/kg fuel)*285.84=6.426 MJ/kg

Total Energy output

=4.042+3.467+6.426 **= 13.935MJ/kg fuel**

Total energy input is 16.78 MJ/kg fuel

Cold gas efficiency of the gasifier

=(13.935/16.78)*100=83.045%

Proposed Energy Audit report

Energy input : 8.26 kWh (already calculated)

Energy output

As per material balance calculation it is found that 1kg of biomass sample produces 0.1544 kmol gas which occupies =0.1544*22.4=3.459 m^3at STP. This implies that vol. of gas produced at 800^0C will be =3. 459 *(1073/273) =13.595 m^3, this is under ideal conditions. But the gas analyser reads product gas as 10 m^3/kg of biomass samples. This difference in yield might be due to the fact that the gasifier is not 100% efficient for which all the biomass are not gasified completely and some losses of product gas might be there.

Energy output was calculated using the heating value of the flue gases.

Flow rate of the flue gases from the gasifier unit 1.0 m^3/ hr for 1kg of fuel as feed rate is 10kg/hr, flow rate of gas can be said as 10 m^3/ hr.

Composition of the flue gas for air gasification is as follows

	O_2	CO	CO_2	CH_4	H_2	N_2
Mole or, Vol%	**13.83**	**9.25**	**8.0**	**2.52**	**14.56**	**49.24**
Flow rate(m^3/hr)	**1.383**	**0.925**	**0.80**	**0.252**	**1.456**	**4.924**

Temperature of the flue gases coming out is 800^0C

The heat energy processed by the flue gases was sapped out before calculating the net heating value of the flue gases.

The heat energy extracted from the flue gases:(density of flue gas =0.345kg/$m^3$and mean specific heat at 900^0C is 3.735KJ/kg.^{0}K)

Heat Energy:$\sum mc_p$*(800-35) = (10*0.345)*3.735*765=12.886 *765= 9857.59875 kJ/kg

= 9.858 MJ/hr=2.738243781 kWhr/hr=**2.74 kW(approx.)**

Energy expected to be obtained from the flue gases will be

Compound	Heating Value, (KJ/m^3)
CO	4074
CH_4	11605
H_2	3450

Output energy expected to be obtained from the different components of the flue gas:

(0.925*4074) + (0.252*11605) + (1.456*3450) = 11716.11 kJ/hr = 11.72 MJ/hr

= 3.2545 kWhr/hr= 3.255 kW (**approx.)**

Total Energy output = 2.74+ 3.255= 5.995kWh per kg of fuel

Net energy output=5.995*10 – 8.26 = 4.3283 kW=51.69 kWh

10 kg of feed (dry basis) produces51.7kWh energy equivalent of synthesis gas and 10 $Nm^3$of gas per hour.

Typical Operation Yields and Efficiencies

The operation results give the following yields and efficiencies:

1 kg of feed (dry basis) yielded 0.4328kWhr equivalent of synthesis Gas.

1 kg of feed yielded 1.0 Nm^3 of gas with 14.56% H_2, 2.52% CH_4,9.25% CO, 8.0% CO_2 and 13.83% $O_2$and 49.24% N_2.

Overall gasification efficiency was 50% (considering LHV of feed to LHV of GasOutput).

(E)Energy Balance and Mass Balance Calculationsfor Saw Dust

Data:

Biomass Fuel - 1 kg/hr (0.912 kg/hr dry)

Steam – 0.5 kg/kg fuel

Moisture is 8.8%

Composition of fuel on %dry basis

H_2 =5.32, C = 45.78, N_2 =0.16, O_2 = 48.65, S = 0.07

Product gas:

H_2 =16.06%, CH_4 = 2.25%, CO = 9.25%, CO_2 = 7.6%, N_2 = 48.84% and O_2=15.25%

HHV=16.2MJ/kg

From experiment air flow rate is noted to be = 0.00767m^3/hr

Specific humidity of supply air be 0.01kg moisture per kg dry air

---

Basis: 1 kg of fuel with 0.088kg moisture, 0.912kg dry fuel

Air: Flow rate = 0.00767m^3/hr (corresponding to Equivalence Ratio=0.2) for Feed rate of 10kg/hr

Thus air flow rate for 1kg/hr of feed is calculated as = 2.76kg/kg of fuel

Steam: 0.5 kg/kg of fuel (since steam rate is 5 kg/hr for 10kg/hr of feed)

Air contains 76.9% N_2 by mass

So N_2 from air

= 0.769* 2.76 = 2.12244 kg N_2/kg fuel

Total N_2 supplied by air and fuel which carry 0.16% N_2

= 2.12244+ 0.0016 = 2.12404 kg N_2 /kg fuel

= 0.07586 kg mol/ kg fuel

We know product gas contains 48.84% N_2 . Hence the amount of product gas produced

= 0.07586/0.4884 = 0.155321 kg mol gas / kg fuel

Product Gas=0.155 kg mol / kg fuel

Oxygen inflow to Gasifier :

Oxygen supplied to gasifier (with air) is

= 0.2315 * 2.76 = 0.6389 kg/kg fuel

Steam supplied = 0.5 kg /kg fuel

So oxygen associated with steam supply

= 16/18 *0.5 = 0.4444 kg /kg fuel

Moisture in fuel = 8.8%

Oxygen with moisture in fuel

= 0.088 * 16/18 = 0.0782222 kg/kg fuel

Oxygen with moisture in air supply

= 0.01 * 2.76 * 16/18 = 0.024533 kg/kg fuel

Oxygen in fuel =48.65% =0.4865kg/kg fuel

Total oxygen flow to gasifier with air steam, moisture in fuel and air

= 0.63894 + 0.44444 + 0.078222 + 0.024333 + 0.4865

= 1.672435 kg/kg fuel

≈ **1.672 kg/kg fuel**

H_2 Balance:

Total H_2 in flow to the gasifier with fuel steam, moisture in fuel and moisture in air

= 0.0532 * 0.912 + 0.5 * 2/18 + 0.088 * 2/18 + 0.01* 2.76*2/18

= 0.11692 kg/kg fuel

H_2 associated with H_2 & CH_4 in dry product gas

= (0.1606 + 0.0225 * 2) * 0.155321

= 0.031934 kg mol/kg fuel

=0.064 kg H_2 /kg fuel

Assuming all S converted to H_2S and removed by the gas cleaning system, hydrogen associated with H_2S in the raw product gas

=0.0007 * 2/32 = 4.375 * 10^{-5} kg/kg fuel

Total hydrogen in the uncleaned dry product gas including that in H_2S

=0.06387 + 4.375 * 10^{-5}= 0.063914 kg/kg fuel

= 0.064 kg H_2 /kg fuel

To find the moisture in the product gas we deduct the hydrogen in dry gas from the total hydrogen in flow obtained earlier using the hydrogen balance.

= Hydrogen in flow – Hydrogen in dry gas

= 0.11692– 0.063914

= 0.053006 ≈ 0.053 kg/kg fuel

Steam associated with this hydrogen in the gas

=0.0533 * 18/2 = 0.477 = **0.477 kg /kg fuel**

<u>Oxygen Balance:</u>

Oxygen associated with CO, CO_2 in dry gas which have half a mol and 1 mol of oxygen respectively.

= (0.5 * 0.0925 + 0.076+0.1525) * 0.155321

=0.27475*0.155321

= 0.04267444 kg mol /kg fuel

= 0.04267444 * 32 = 1.3655822 kg /kg fuel

Oxygen associated with the steam in gas

= 0.477 * 16/18 =0.417777 kg /kg fuel

Total Oxygen in gas = 1.3655822 + 0.4177777 = =1.783359978 = **1.783 kg/kg fuel**

Here we note that this is slightly more than the oxygen in flow of **1.672 kg /kg fuel** calculated earlier. This must be due to measurement errors in the given data on fuel gas composition.

Carbon Balance:

Total carbon associated with CO, CO_2 and CH_4 in dry gas whose production rate has been computed earlier as 0.152 kg-mol / kg fuel is

= (0.0925+0.076 +0.0225) * 0.155321

= 0.191 * 0.155321 = 0.029666311 kg mol/ kg fuel

= 0.029666311 * 12 = 0.355995732**= 0.356kg/kg fuel**

Carbon input is found from the composition of the fuel

= 0.4578**kg /kg fuel**

Carbon conversion efficiency

= (0.356/0.4578) * 100

= 0.7776=77.76%

Energy Analysis

Components	**Heats of Combustion, MJ/kg mol**
CO	**282.99**
H_2	**285.84**
CH_4	**890.36**

Energy output with CO

=0.0925*0.155321(kg mol CO/kg fuel)*282.99=4.065772MJ/kg fuel

Energy output with CH_4

=0.0225*0..155321 (kg molCH_4/kg fuel)*890.36=3.11156MJ/kg fuel

Energy output with H_2

=0.1606*0.155321(kg molH_2/kg fuel)*285.84=7.130151MJ/kg

Total Energy output

=4.065772+3.11156+7.130151**=14.307MJ/kg fuel**

Total energy input is 16.2 MJ/kg fuel

Cold gas efficiency of the gasifier

=(14.3075/16.2)*100= 88.32%

Proposed Energy Audit report

Energy input: 8.26kWh (already calculated)

Energy output :

As per material balance calculation it is found that 1kg of biomass sample produces 0.1553 kmol gas which occupies =0.1553*22.4=3.48m^3at STP. This implies that vol. of gas produced at 800^{0}C will be =3.48*(1073/273) =13.68 m^3, this is under ideal conditions. But the gas analyser reads product gas as 11 m^3/kg of biomass samples. This difference in yield might be due to the fact that the gasifier is not 100% efficient for which all the biomass are not gasified completely and some losses of product gas might be there.

Energy output was calculated using the heating value of the flue gases.

Flow rate of the flue gases from the gasifier unit 1.1 m^3/ hr for 1kg of fuel as feed rate is 10kg/hr, flow rate of gas can be said as 11 m^3/ hr.

Composition of the flue gas for air gasification is as follows

	O_2	**CO**	**CO_2**	**CH_4**	**H_2**	**N_2**
Mole or, Vol%	**15.25**	**9.25**	**7.6**	**2.25**	**16.06**	**48.84**
Flow rate(m^3/hr)	**1.6775**	**1.0175**	**0.836**	**0.2475**	**1.7666**	**5.3724**

Temperature of the flue gases coming out is 800^{0}C

The heat energy processed by the flue gases was sapped out before calculating the net heating value of the flue gases.

The heat energy extracted from the flue gases:(density of flue gas =0.345kg/$m^3$and mean specific heat at 900^{0}C is 3.735KJ/kg.^{0}K)

Heat Energy:$\sum mc_p*(800-35)$ = (11*0.345)*3.735*765=14.174325 *765= 10843.35863 kJ/kg

= 10.8434MJ/hr=3.012068159kWhr/hr=**3.012 kW(approx.)**

Energy expected to be obtained from the flue gases will be

Compound	Heating Value, (KJ/m^3)
CO	4074
CH_4	11605
H_2	3450

Output energy expected to be obtained from the different components of the flue gas:

(1.0175 *4074) + (0.2475*11605) + (1.7666*3450) = 13112.3025 kJ/hr = 13.112 MJ/hr

= 3.642335 kWhr/hr= 3.64234 kW (**approx.)**

Total Energy output = 3.102+ 3.64234= 6.744 kWh per kg of fuel

Net energy output= 6.74434 *10– 8.26 = 59.183 kWh

10 kg of feed (dry basis) produces 59.1834kWh energy equivalent of synthesis gas and 11 $Nm^3$of gas per hour.

Typical Operation Yields and Efficiencies

The operation results give the following yields and efficiencies:

1 kg of feed (dry basis) yielded 0.508kWhr equivalent of synthesis Gas.

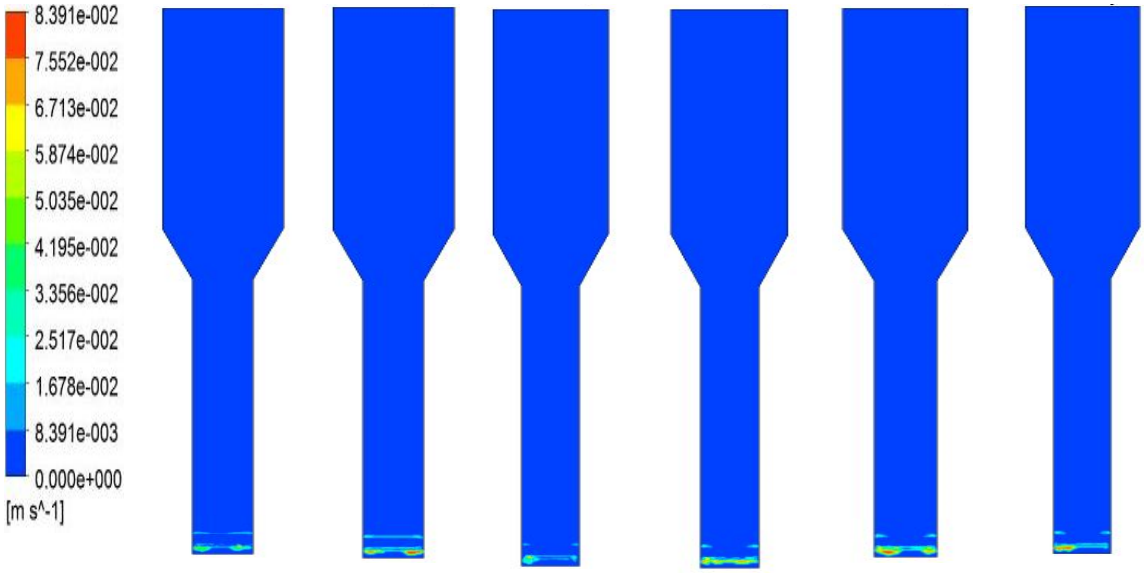
1 kg of feed yielded 1.0 Nm^3 of gas with 16.06% H_2, 2.25% CH_4,9.25% CO, 7.6% CO_2 and 15.25% $O_2$and 48.84% N_2.

Overall gasification efficiency was 50% (considering LHV of feed to LHV of GasOutput).

APPENDIX-III

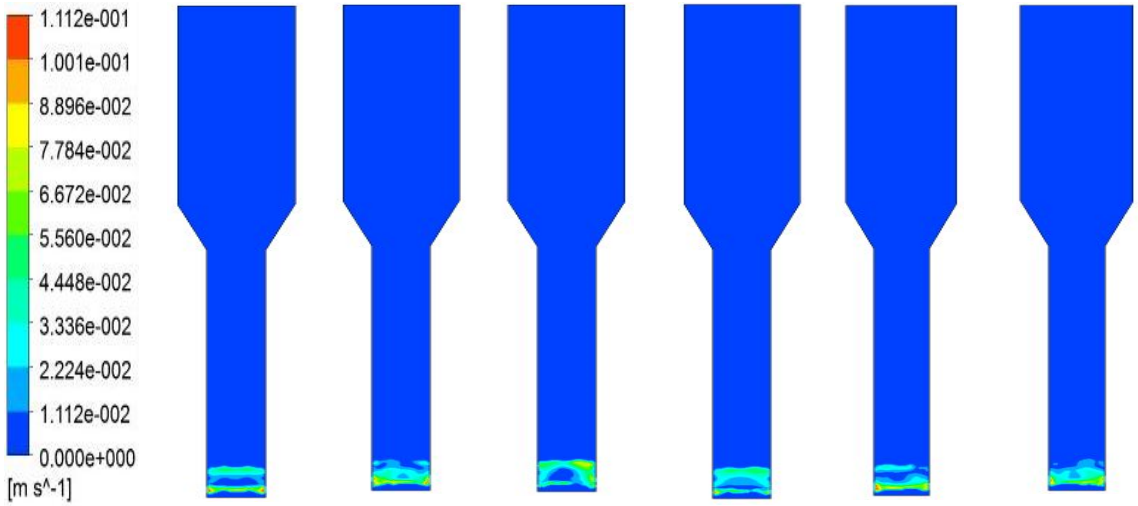
CFD MODELLING RESULTS FOR DIFFERENT BIOMASS SAMPLES

(A) Contour plots of velocity Profile for Sugarcene bagasse at different velocitiesfor initial static bed height of 0.1m



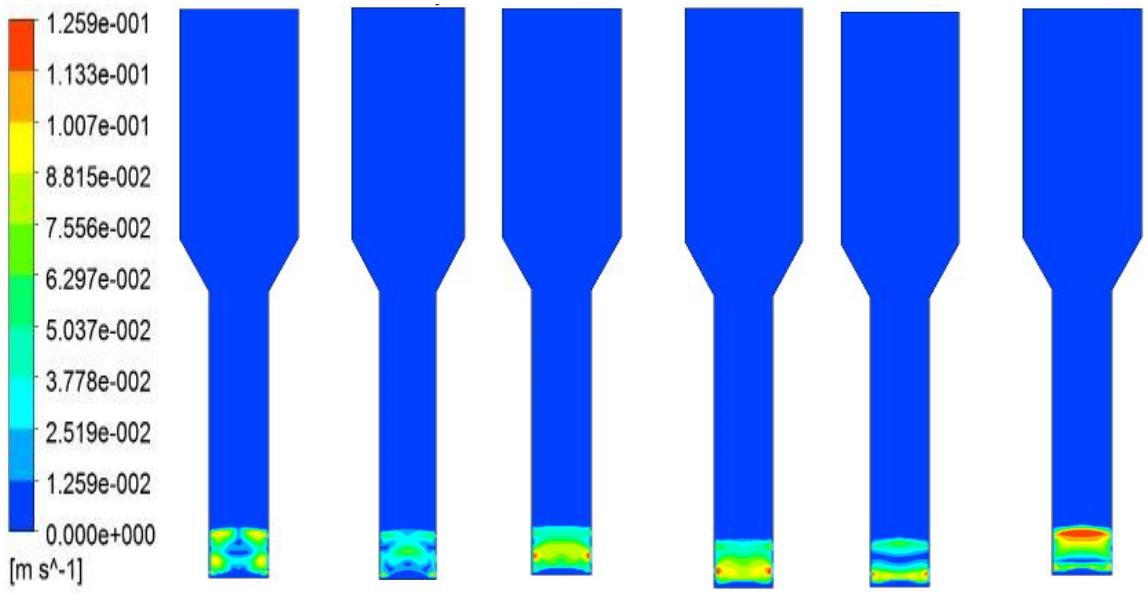
Time 10s 20s 30s 40s 50s 60s

(a) At air velocity = 0.2m/s

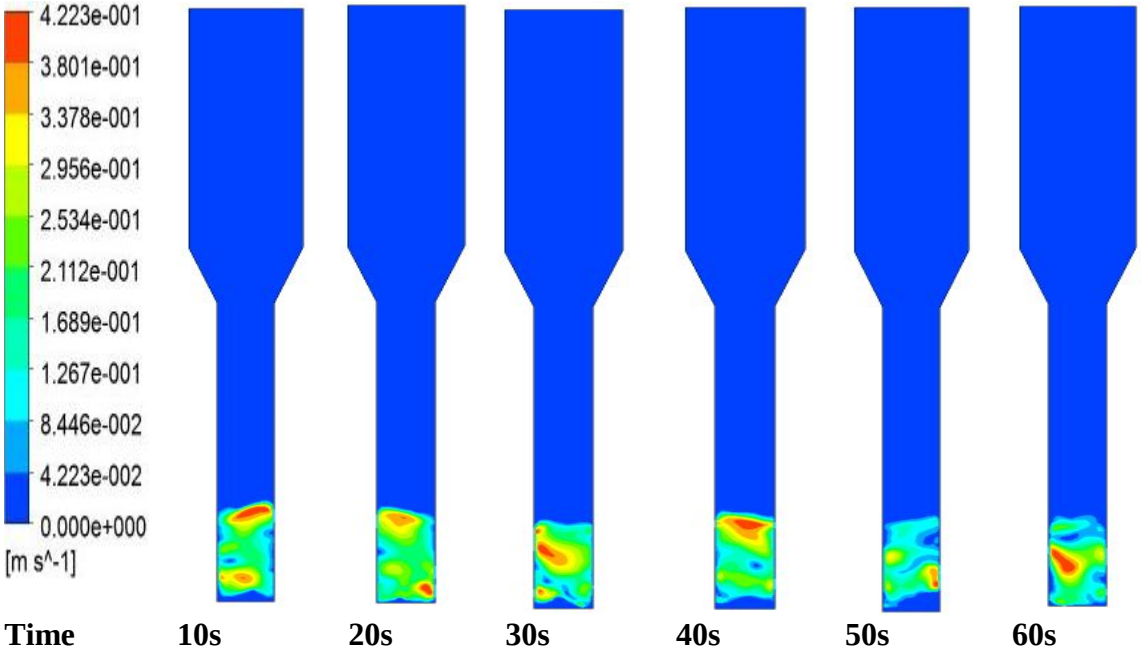


Time 10s 20s 30s 40s 50s 60s

(b) At air velocity = 0.5m/s



(c)At Air Velocity 0.7 m/s.



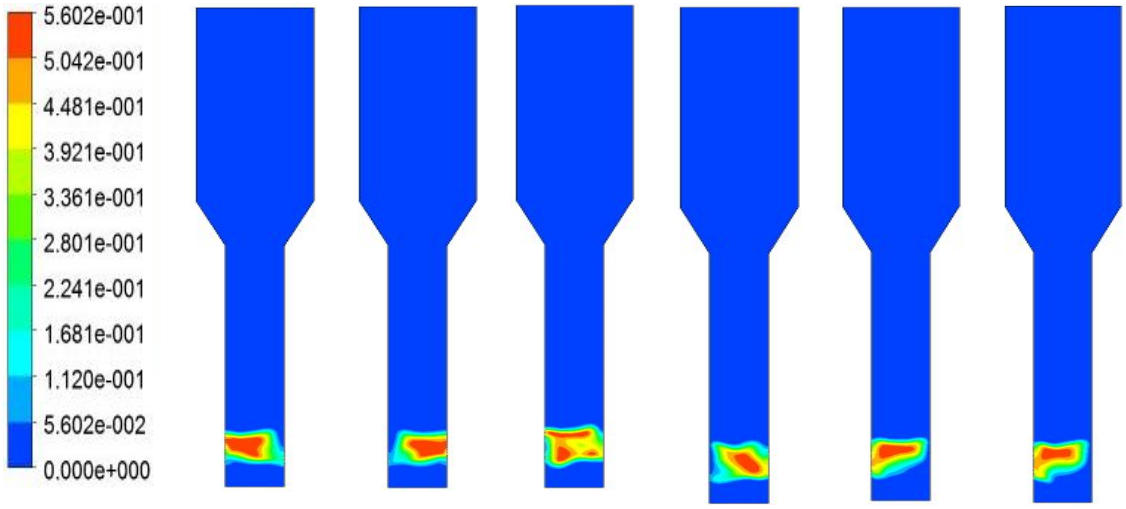
(d) At Air Velocity 0.9 m/s.

Sugarcane bagasse was considered for sample study. From these contour plots velocity profiles it is seen that proper fluidization is observed at 0.9m/s air velocity**.**

Therefore air velocity of 0.9m/s is considered for further studies using both 2D and 3D simulations with all the biomass samples.

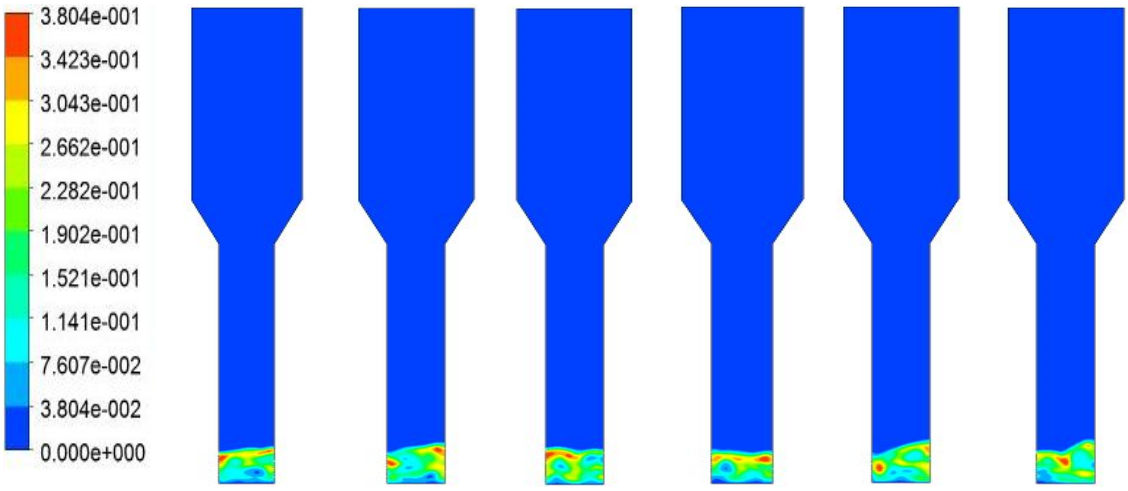
(B) Contour plots of volume fractions for six different biomass samples at 0.9m/s air velocity

I. With 2D-Simulations



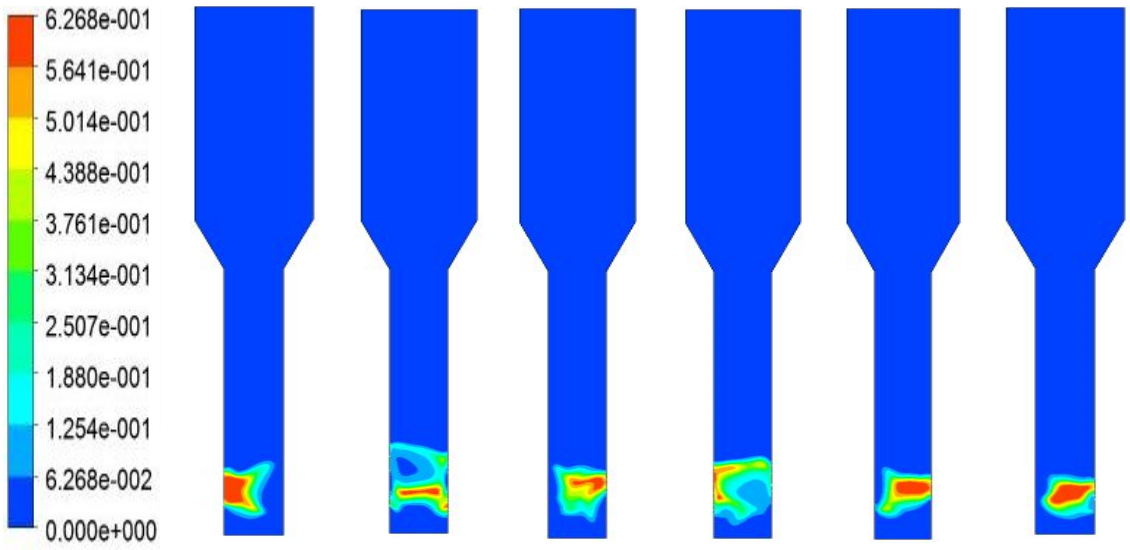
Time 10s 20s 30s 40s 50s 60s

(i)For coconut coir

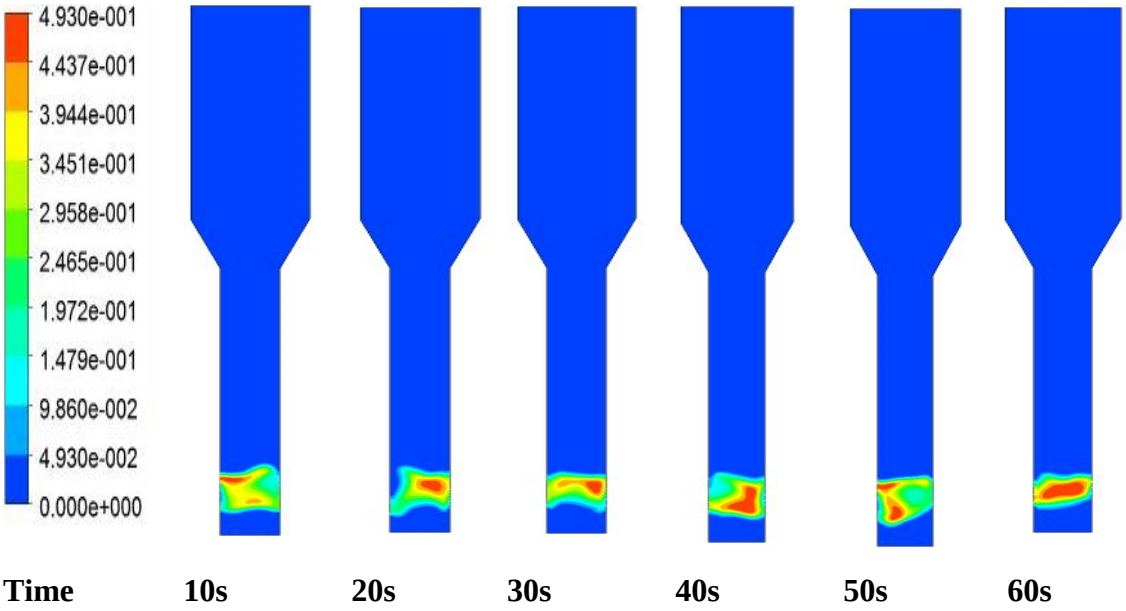


Time 10s 20s 30s 40s 50s 60s

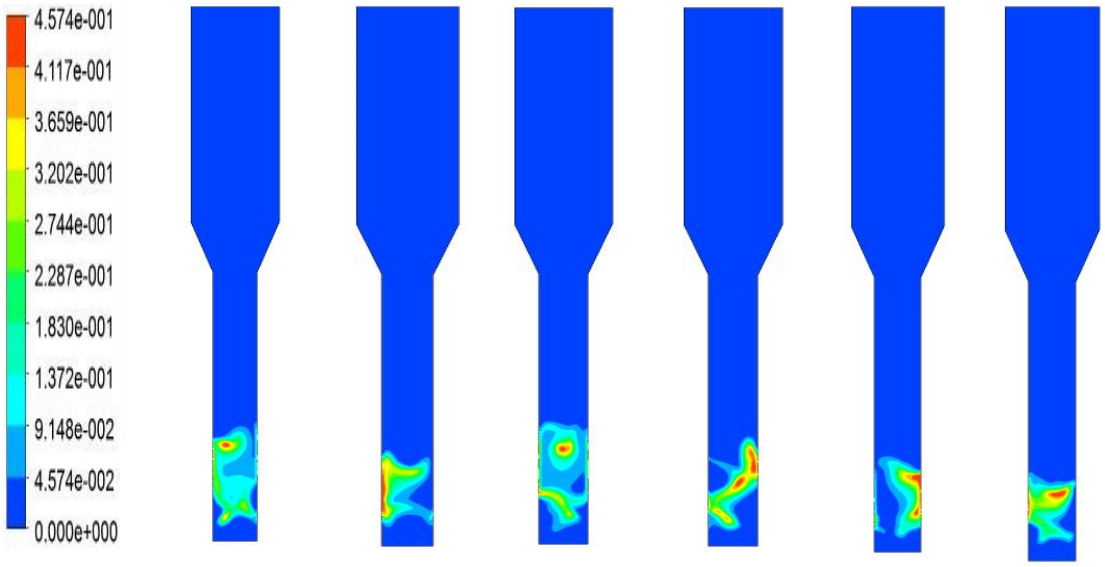
(ii) For wood chips



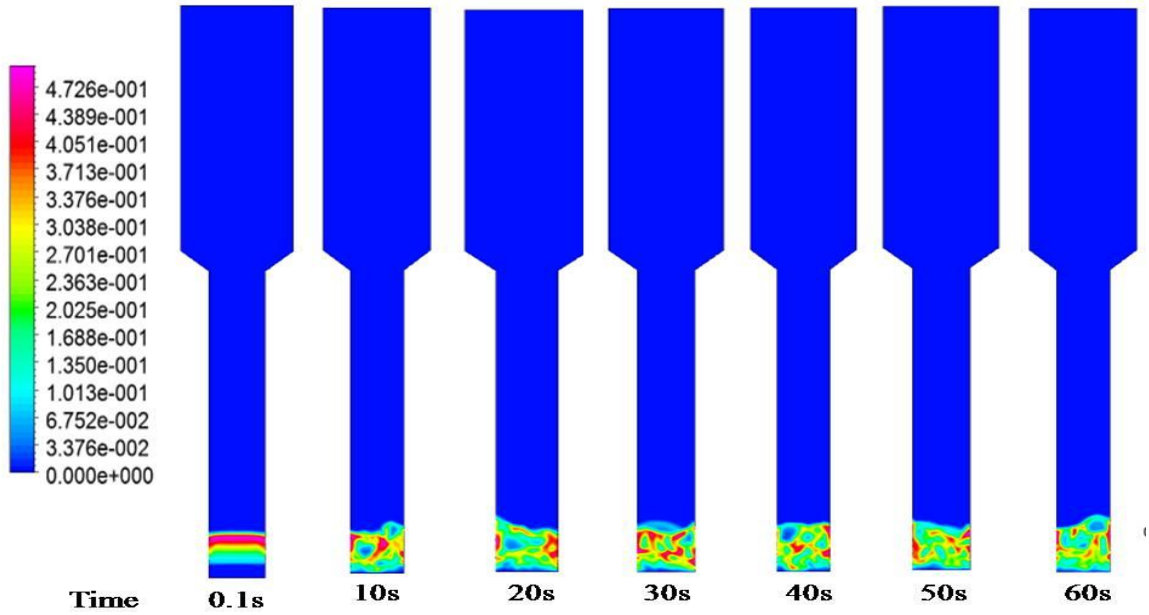
(iii) for rice straw



(iv) For sugarcane bagasse

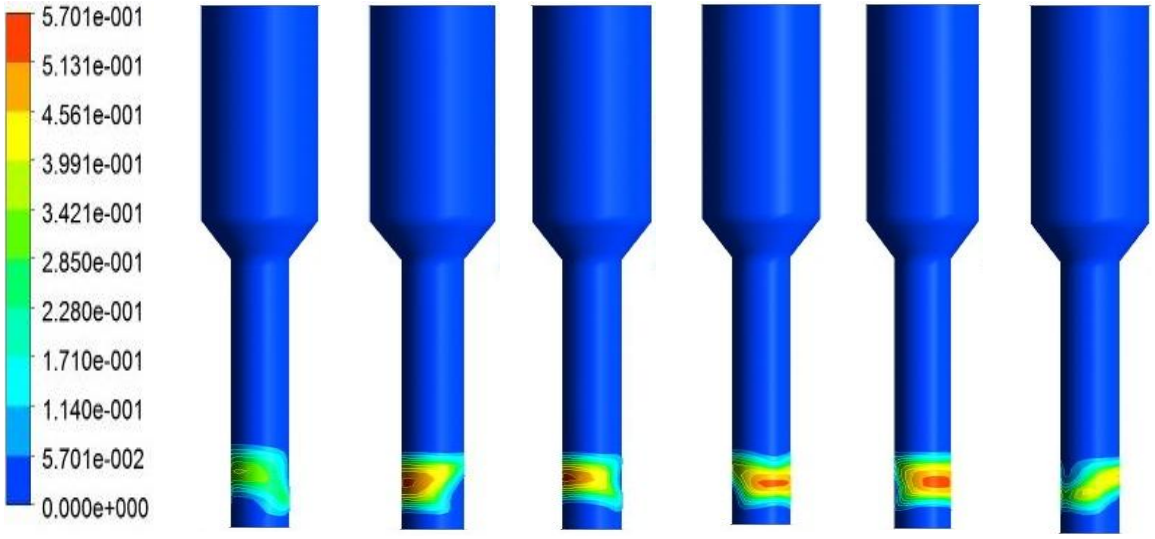


(v) For saw dust

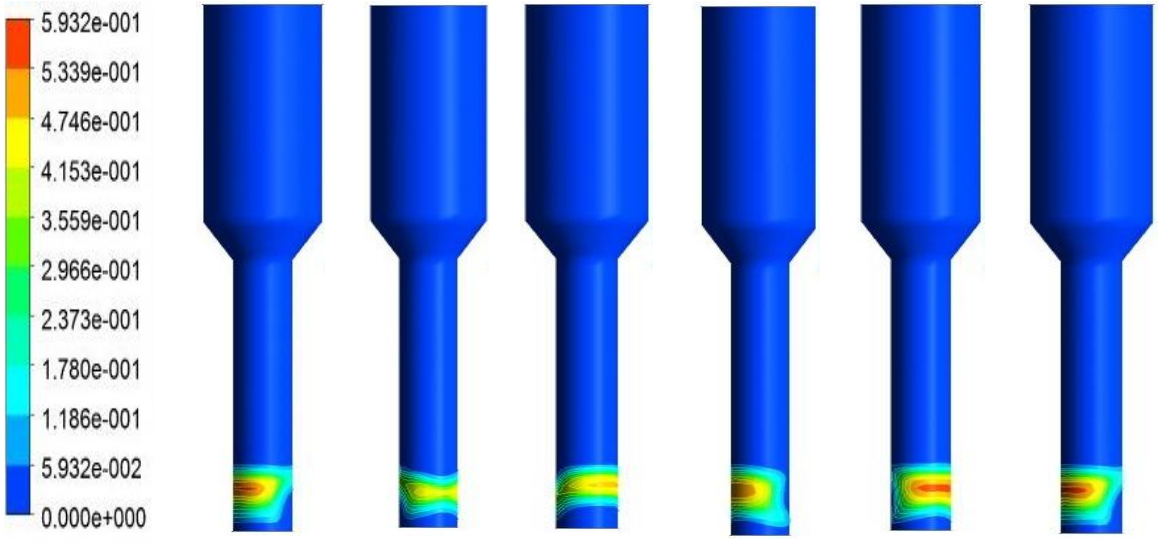


(vi) For rice husk

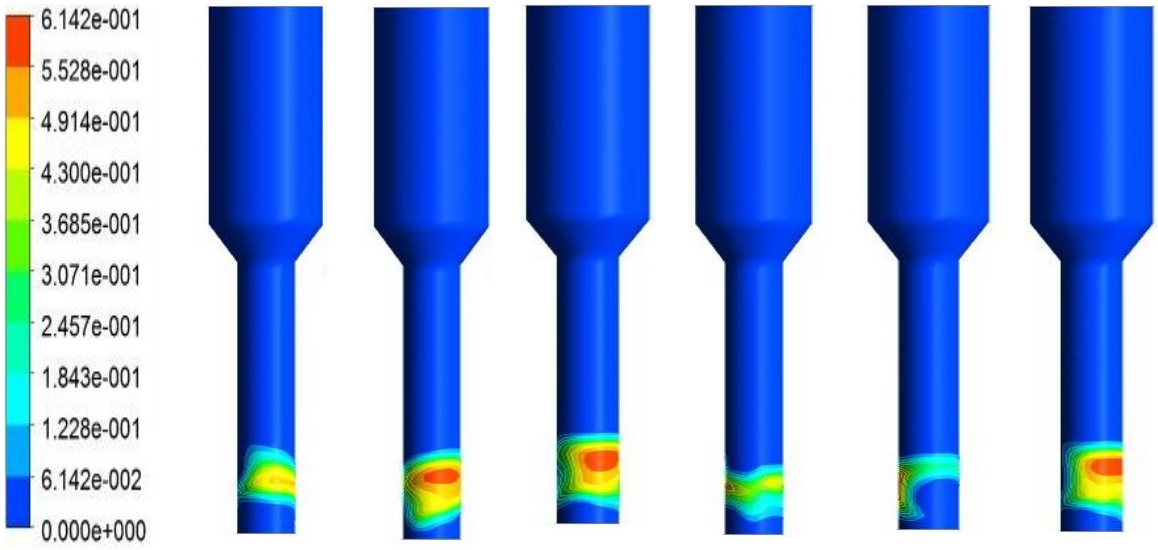
II. With 3D Simulations



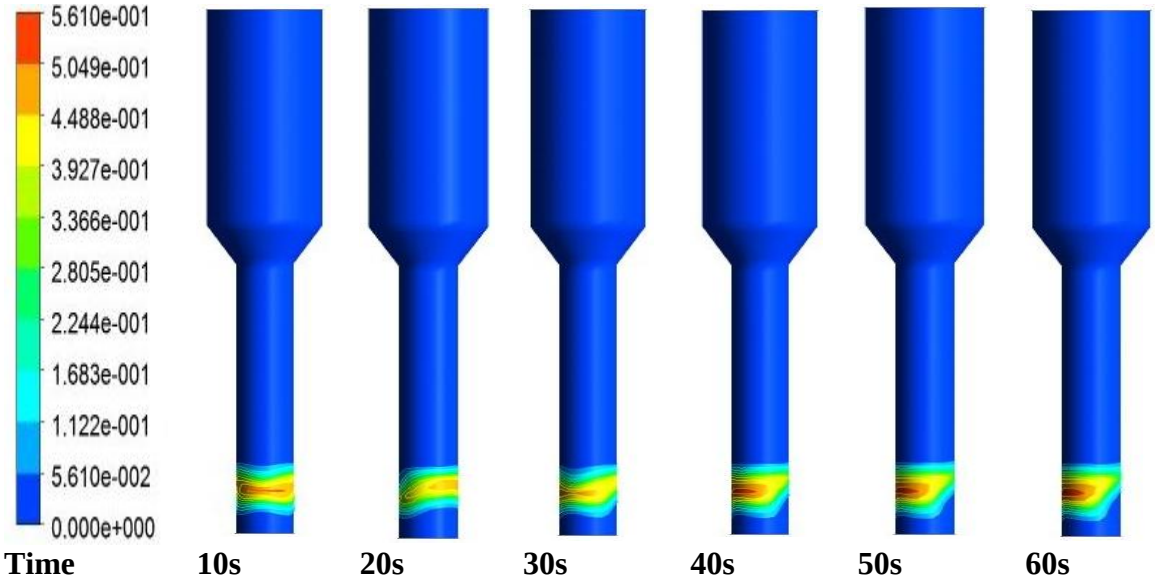
(i) For Coconut coir



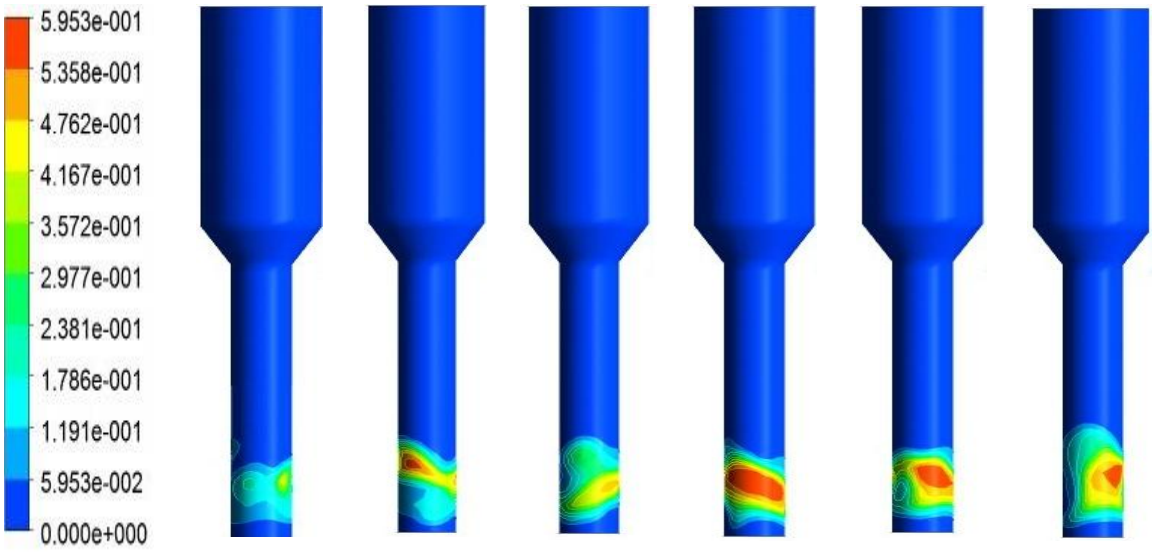
(ii) For wood chips



(iii) For rice straw

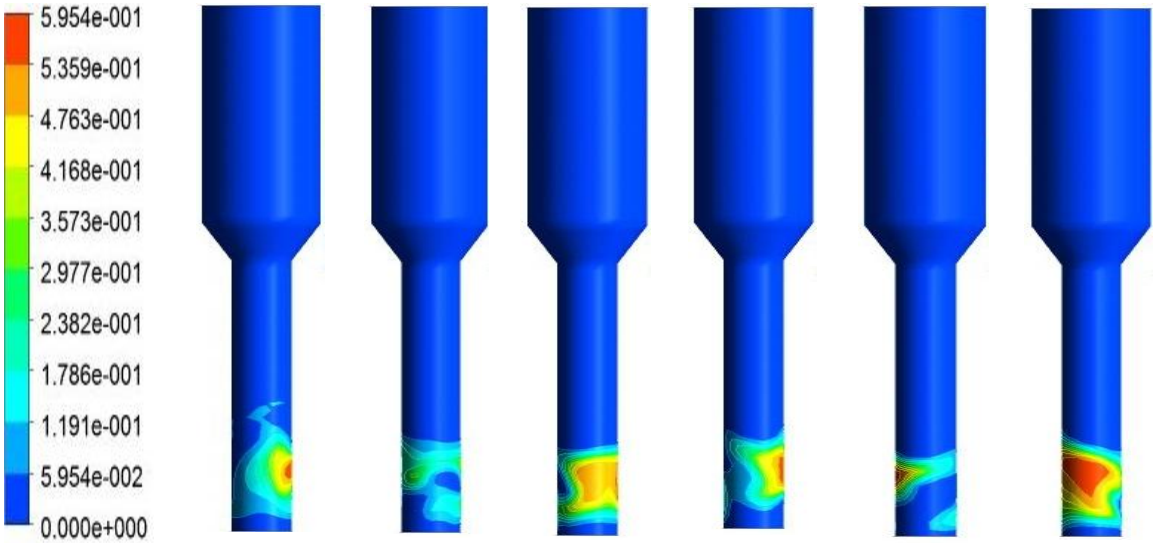


(iv) For sugarcane bagasse



Time 10s 20s 30s 40s 50s 60s

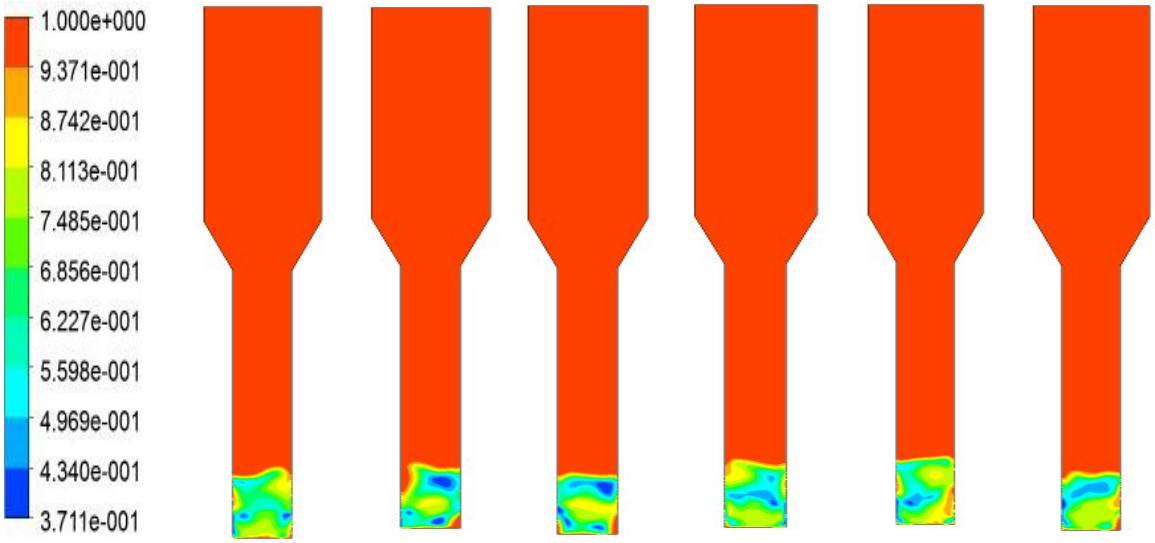
(v) For sawdust



Time 10s 20s 30s 40s 50s 60s

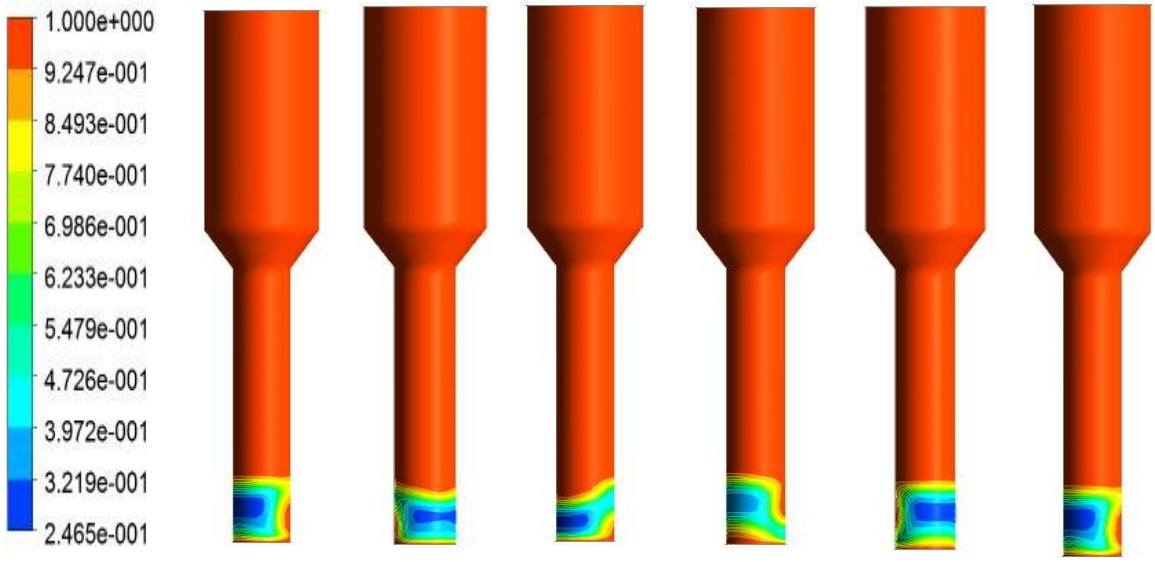
(vi) For rice husk

(C) contour plot for air volume fraction



Time 10s 20s 30s 40s 50s 60s

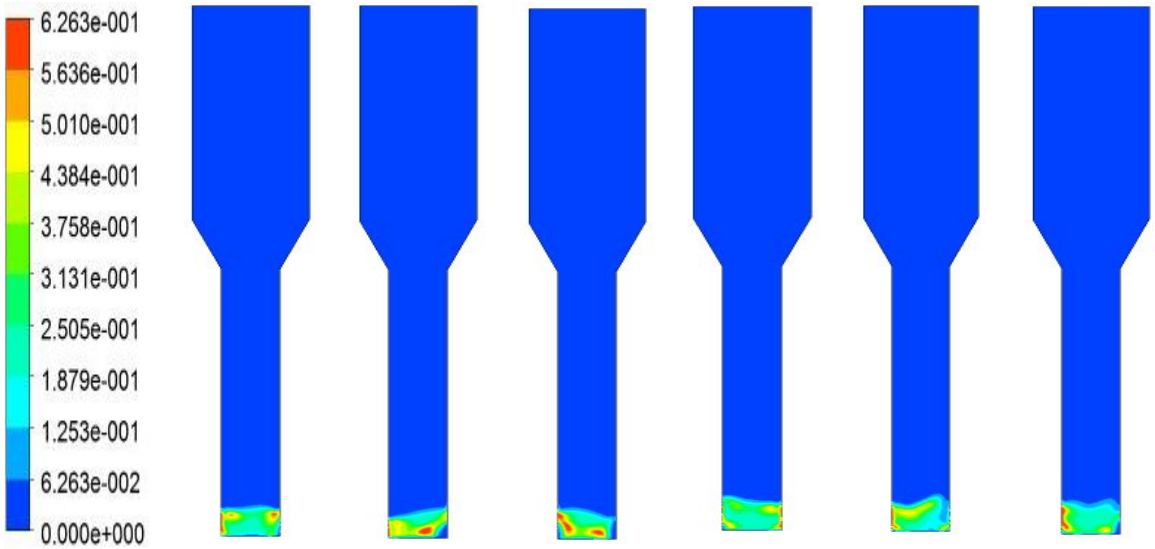
I. With 2D simulations



Time 10s 20s 30s 40s 50s 60s

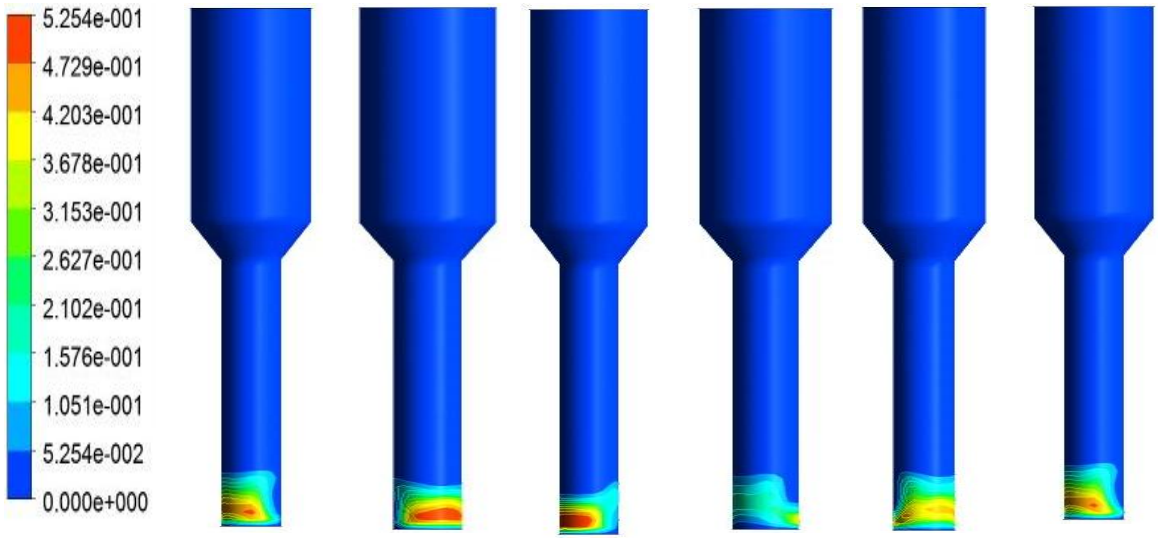
II. With 3D simulations

(D) Contour plots for volume fraction of bed material, sand



Time 10s 20s 30s 40s 50s 60s

I. With 2D simulations

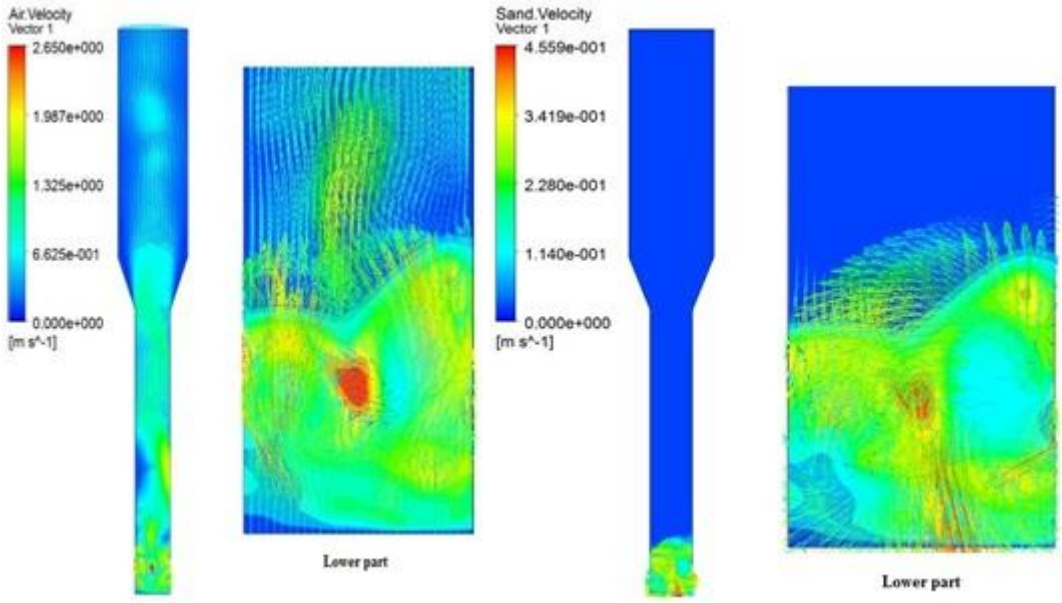


Time 10s 20s 30s 40s 50s 60s

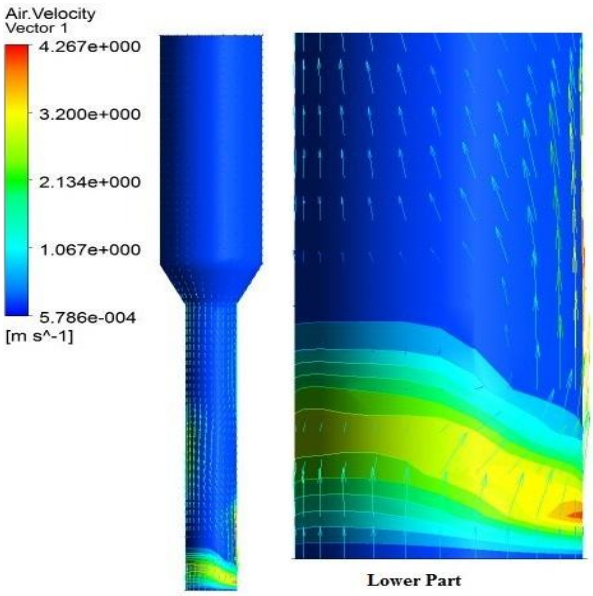
II. With 3D Simulations

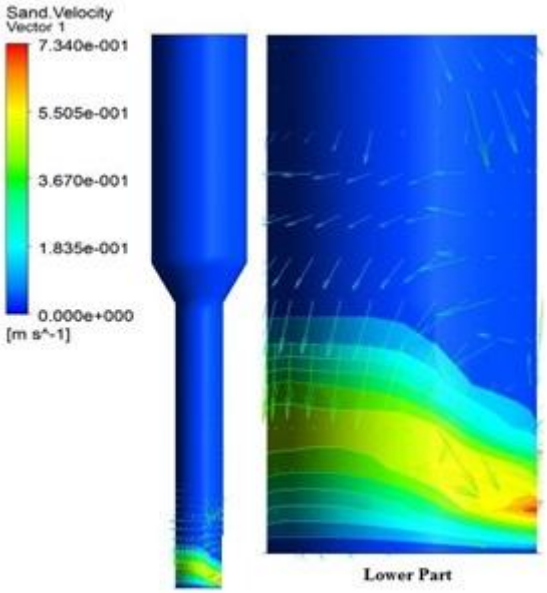
(E) Vector plots for 2D and 3D Simulations

2 D vector plot for Air and Sand

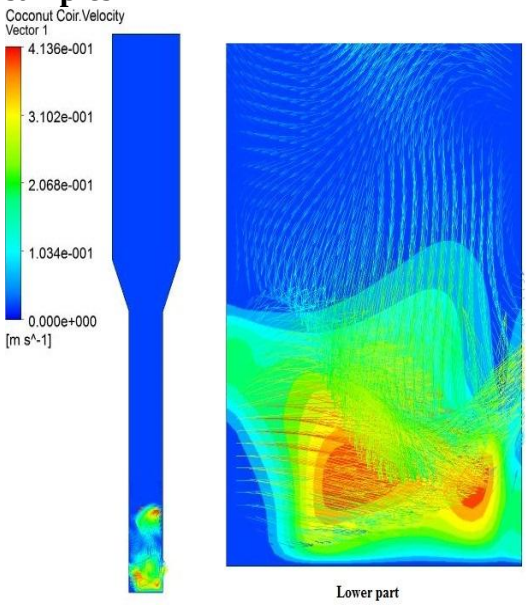


3D Vector plots for Air and Sand

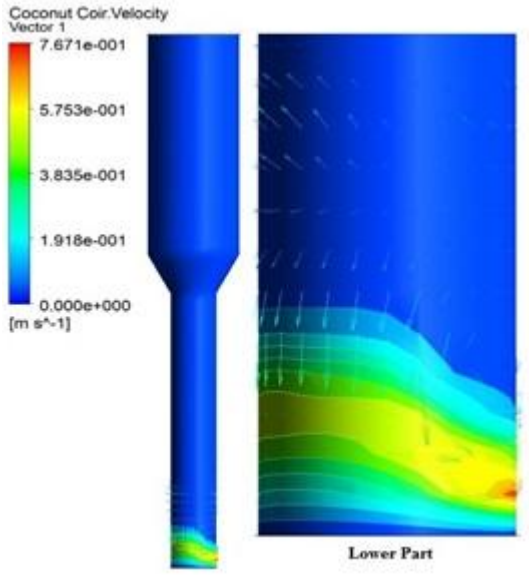


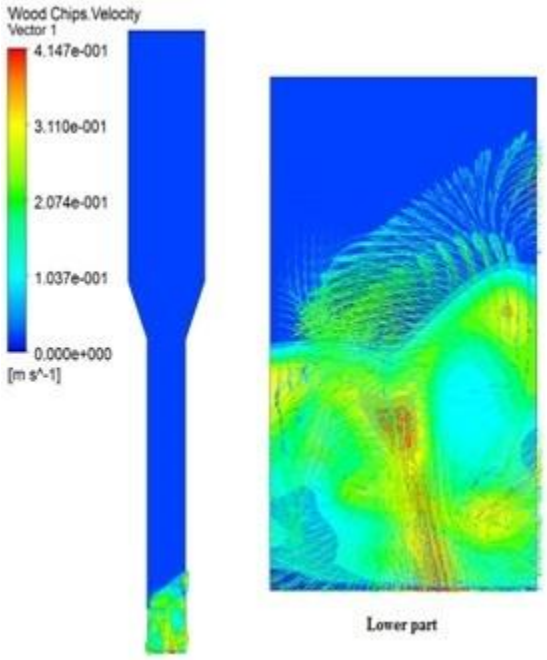


2D vector plots for different biomass samples

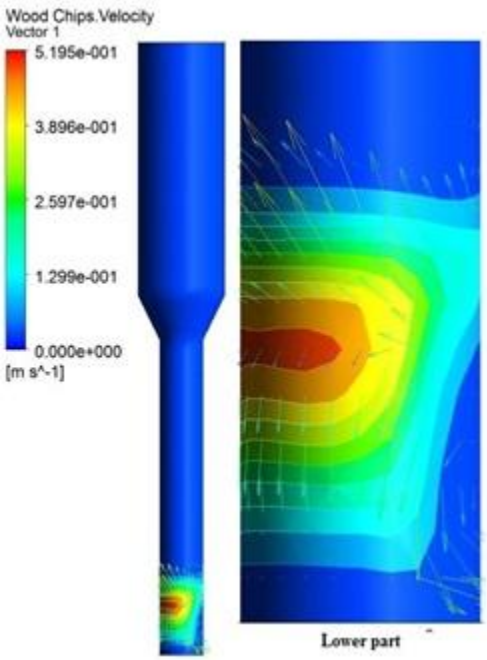


3D vector plots for different biomass samples

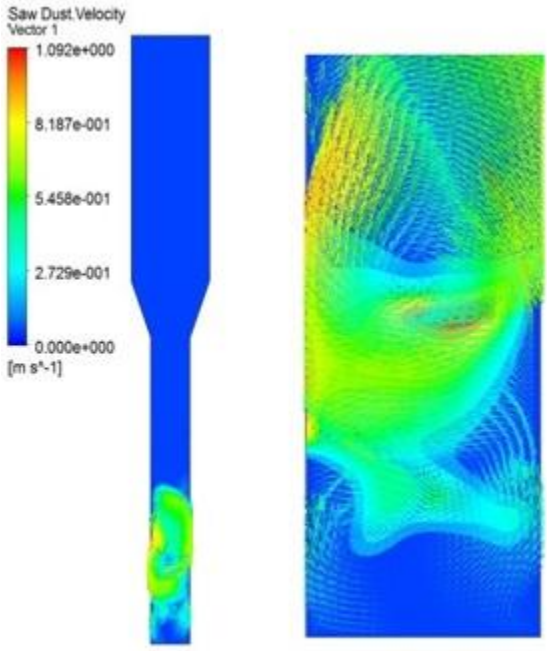




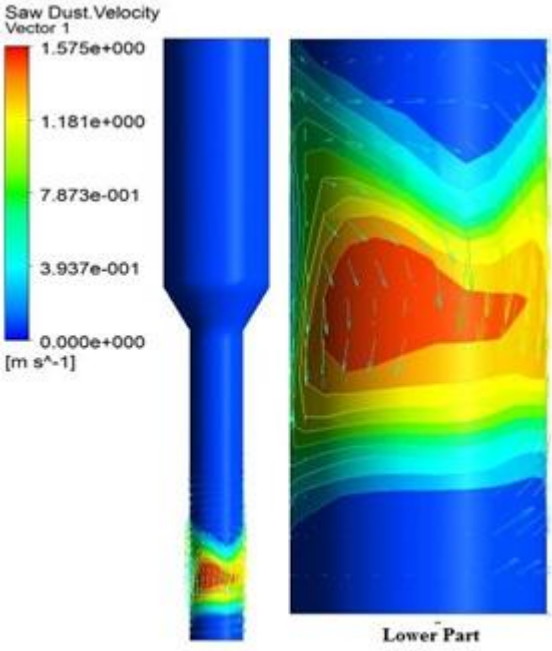
Lower part



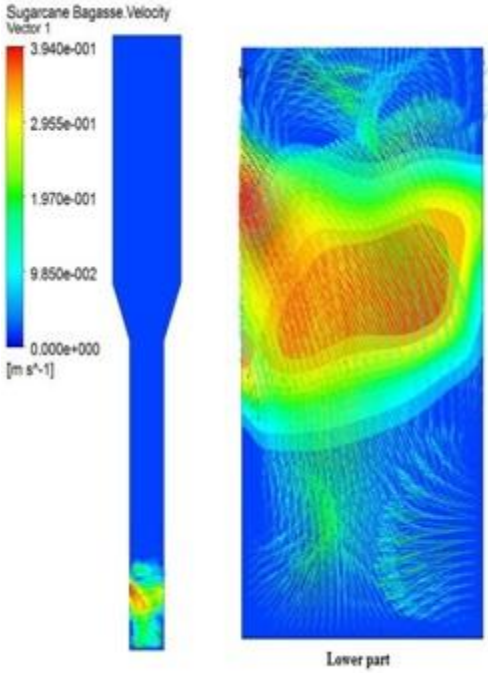
Lower part



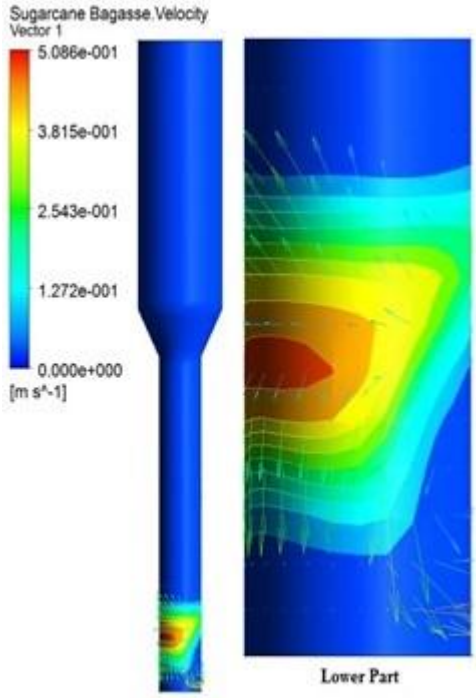
Lower Part



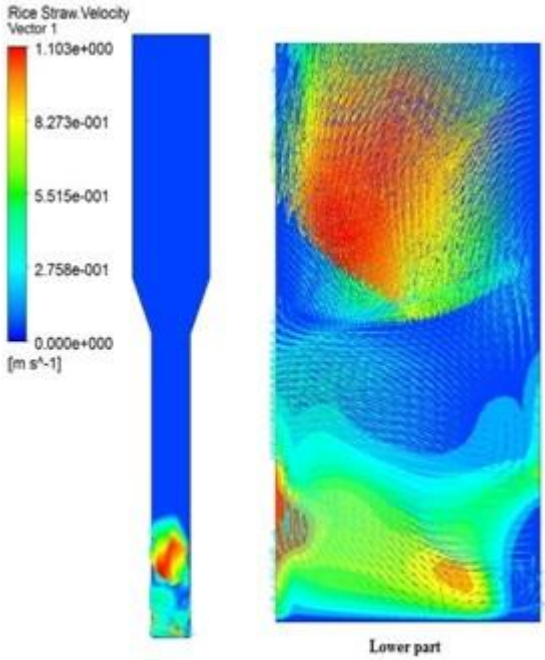
Lower Part



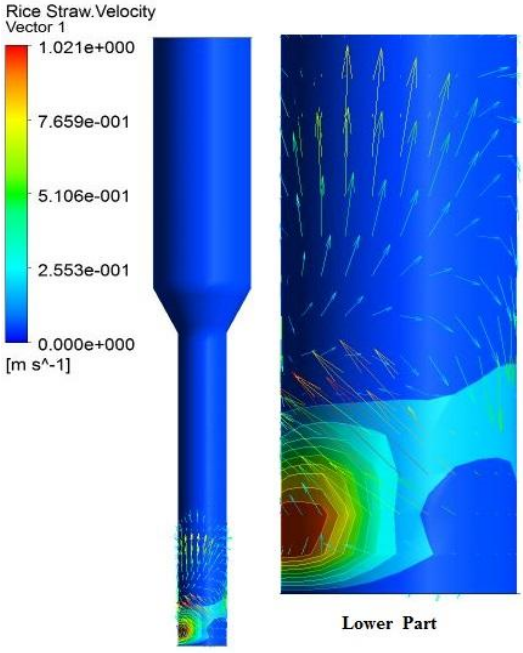
Lower part



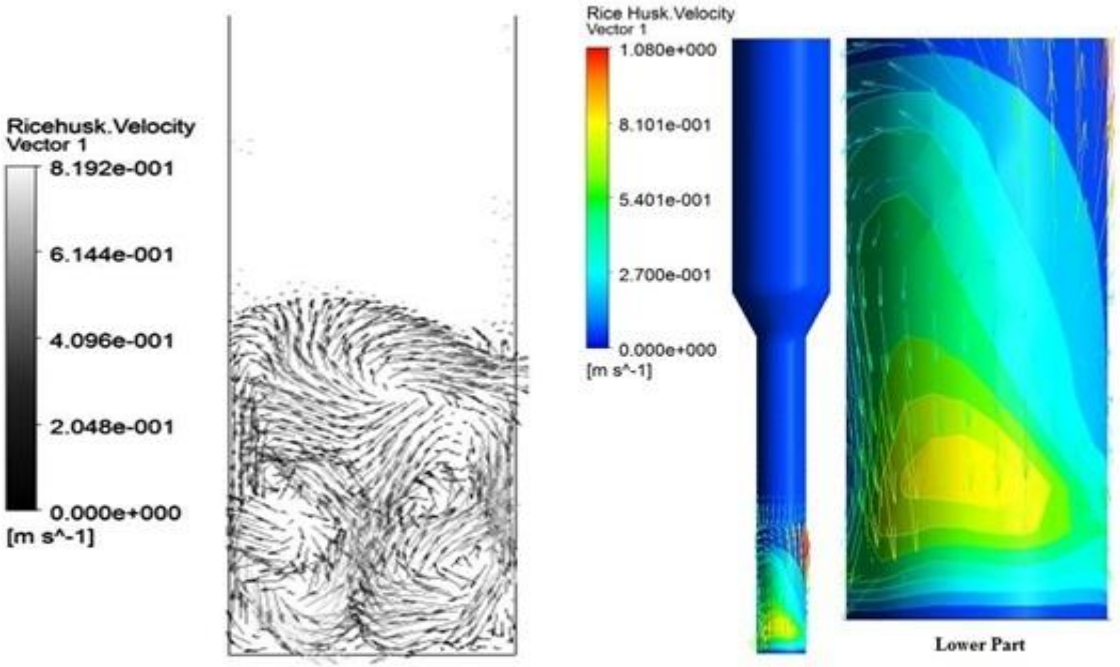
Lower Part



Lower part

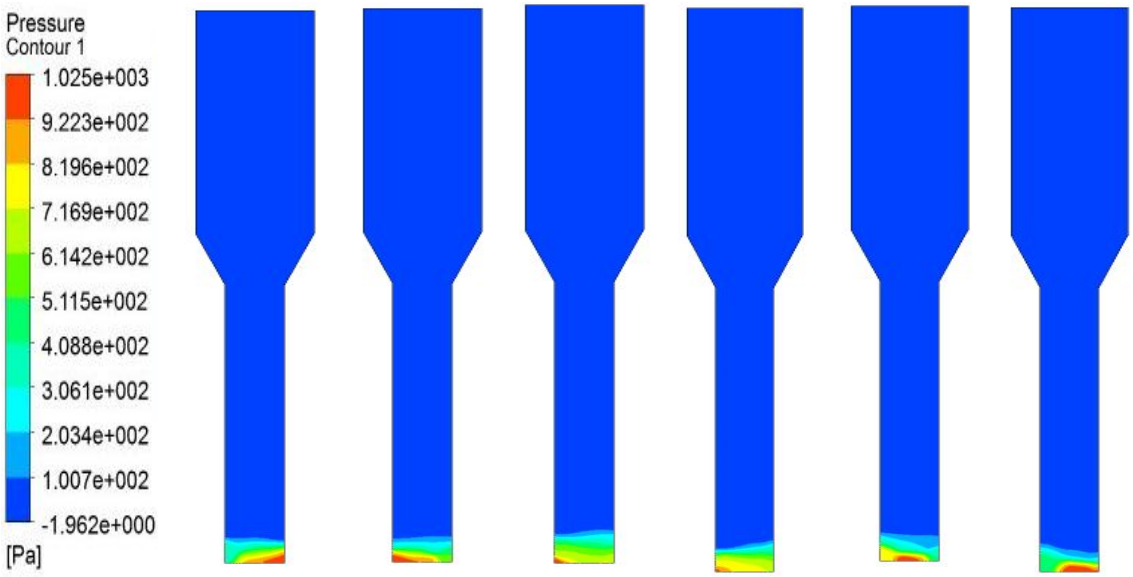


Lower Part

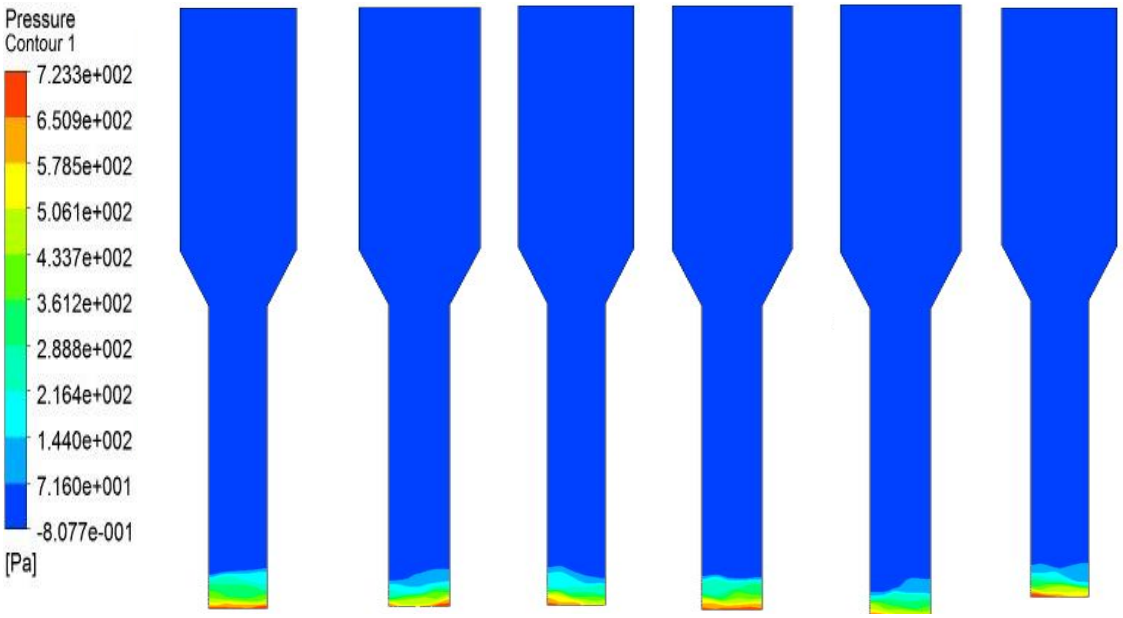


(F) Contour plots for pressure drop profile for different biomass samples

I. With 2D simulations

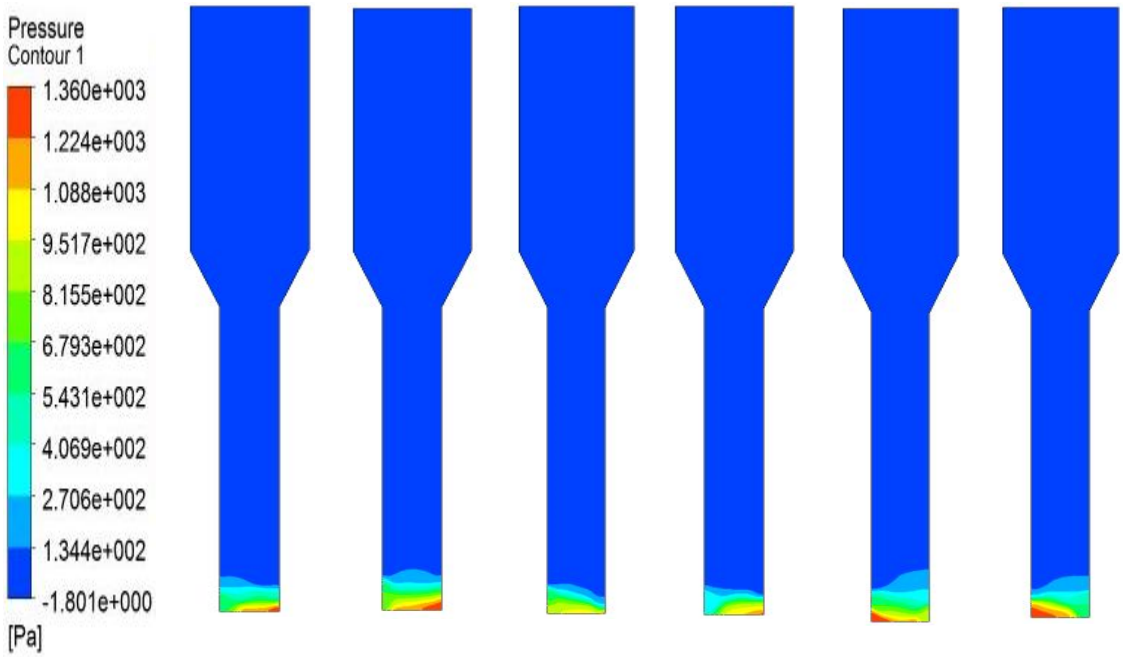


(a) For coconut coir



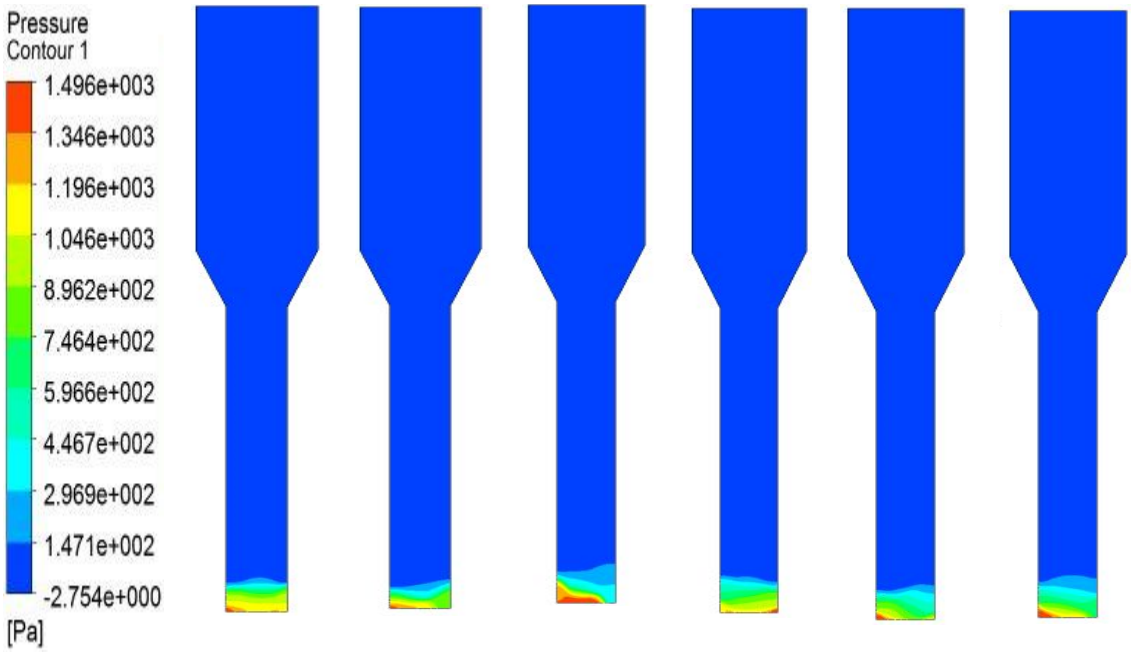
Time 10s 20s 30s 40s 50s 60s

(b) For wood chips



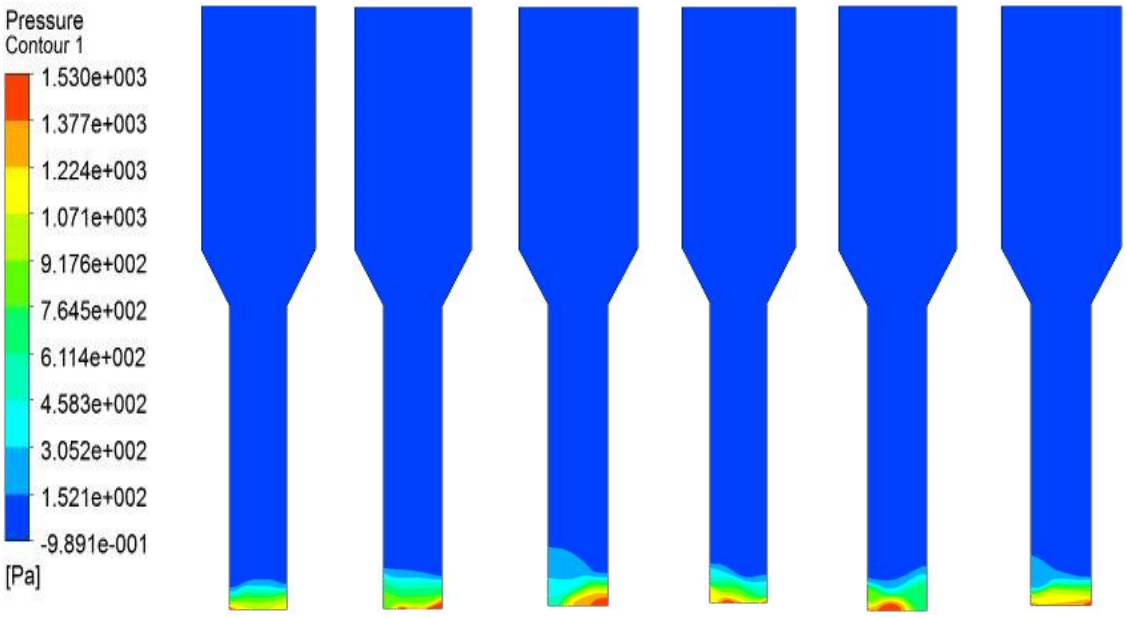
Time 10s 20s 30s 40s 50s 60s

(c) For rice straw



Time 10s 20s 30s 40s 50s 60s

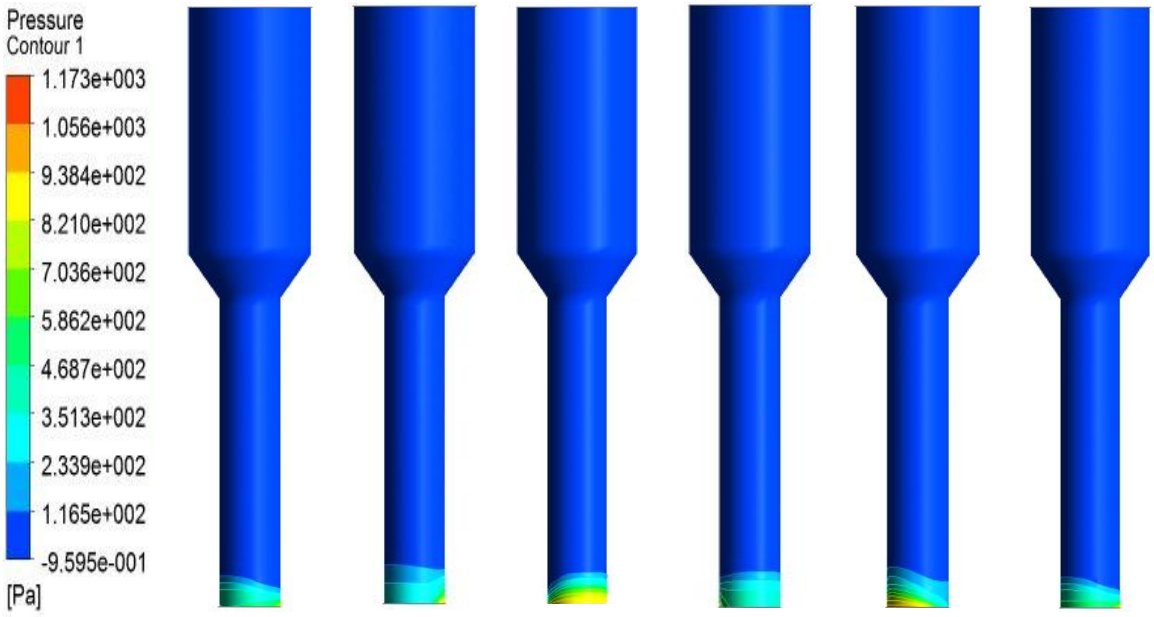
(d) For sugarcane bagasse



Time 10s 20s 30s 40s 50s 60s

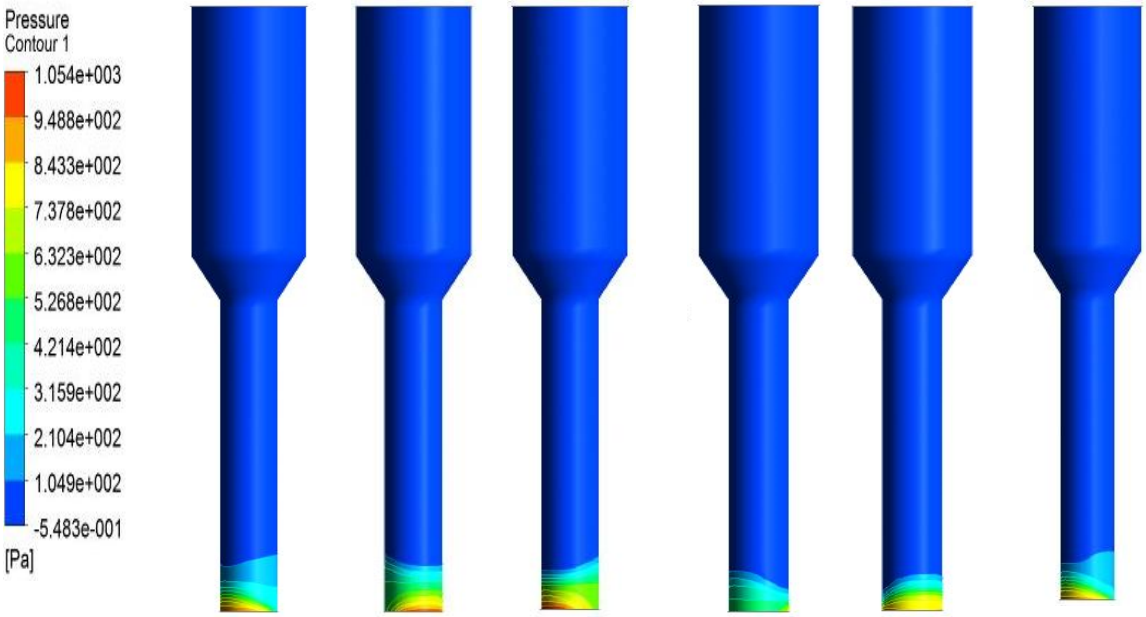
(e) For saw dust

II. With 3D Simulations



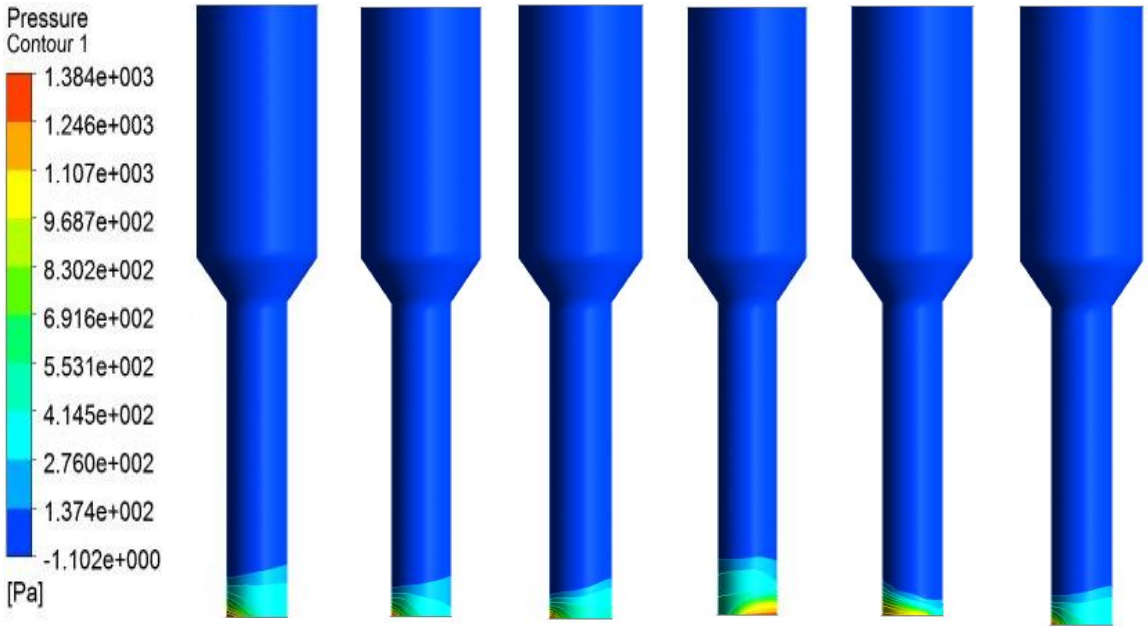
Time 10s 20s 30s 40s 50s 60s

(i) For coconut coir



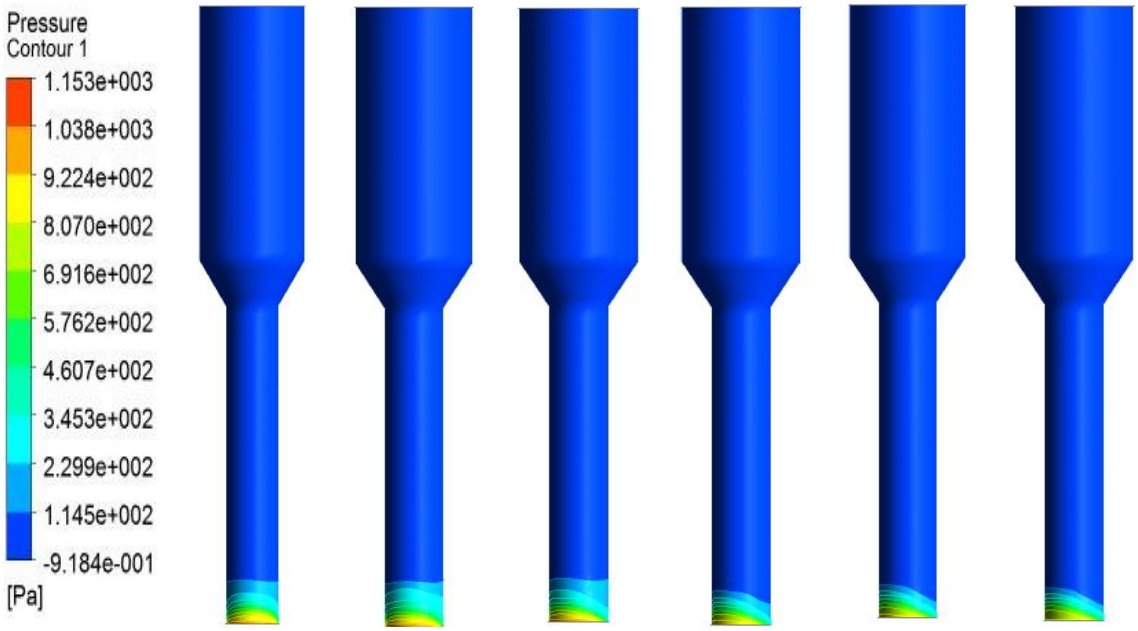
Time 10s 20s 30s 40s 50s 60s

(ii) For wood chips



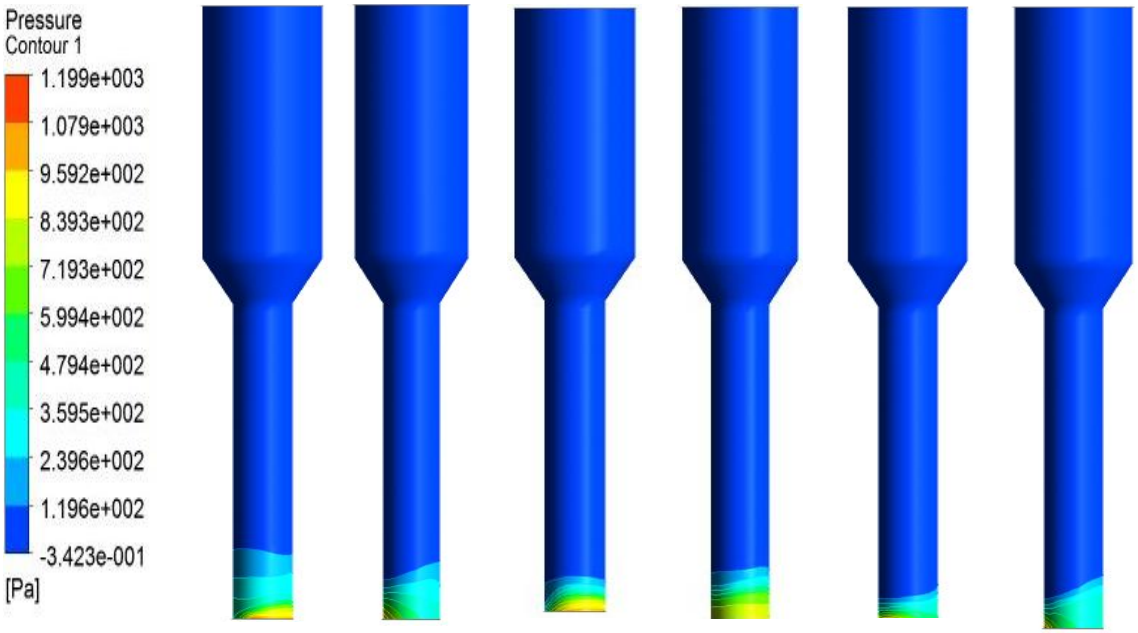
Time **10s** **20s** **30s** **40s** **50s** **60s**

(iii) For rice straw



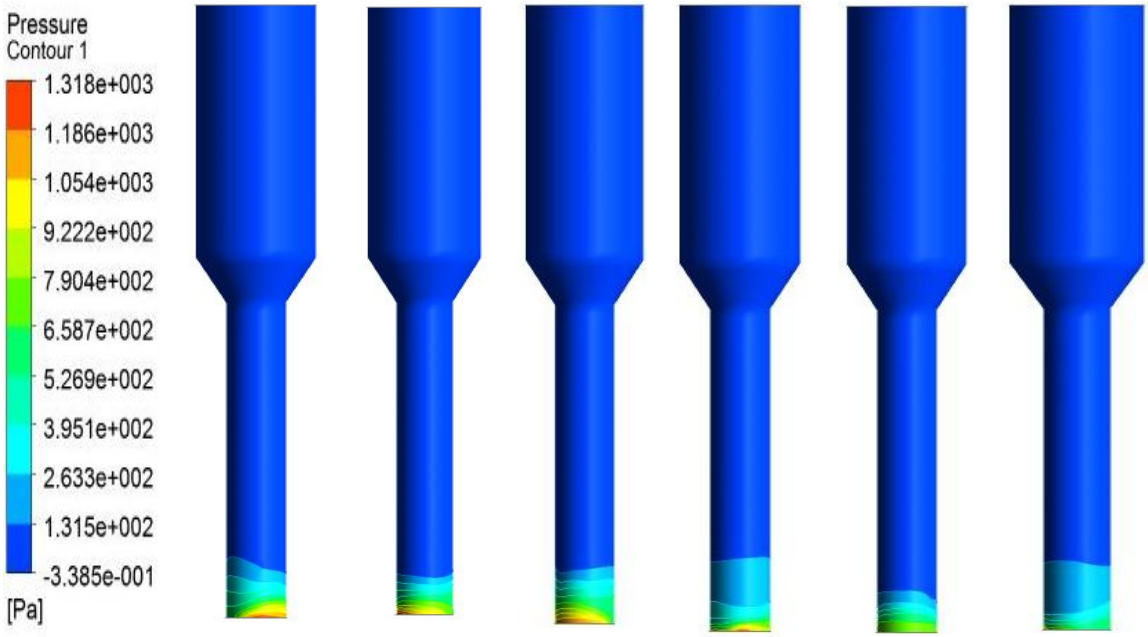
Time **10s** **20s** **30s** **40s** **50s** **60s**

(iv) For sugarcane bagasse



Time 10s 20s 30s 40s 50s 60s

(v) For saw dust

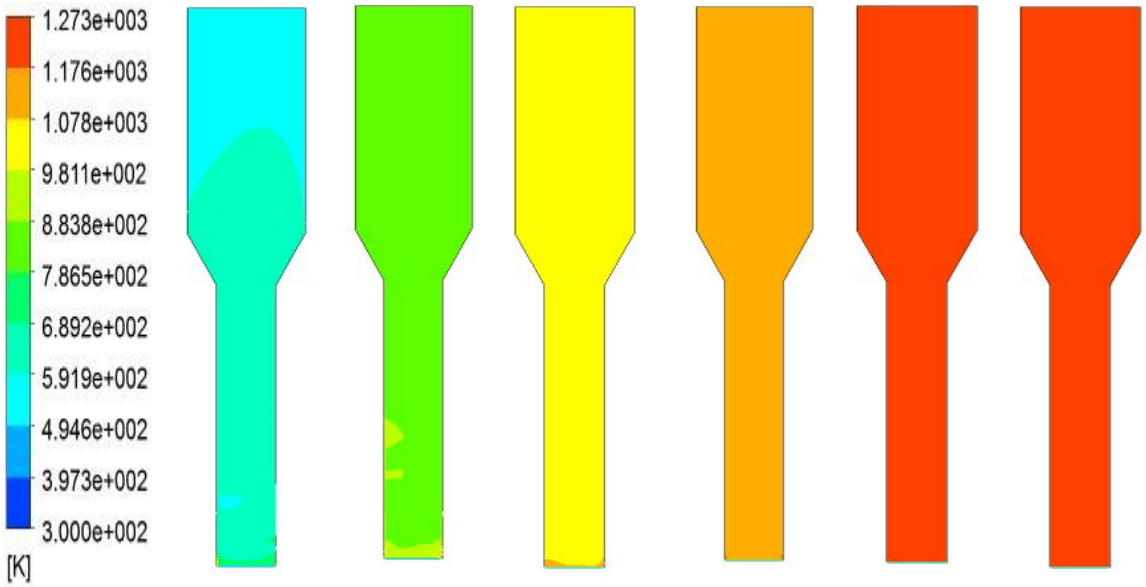


Time 10s 20s 30s 40s 50s 60s

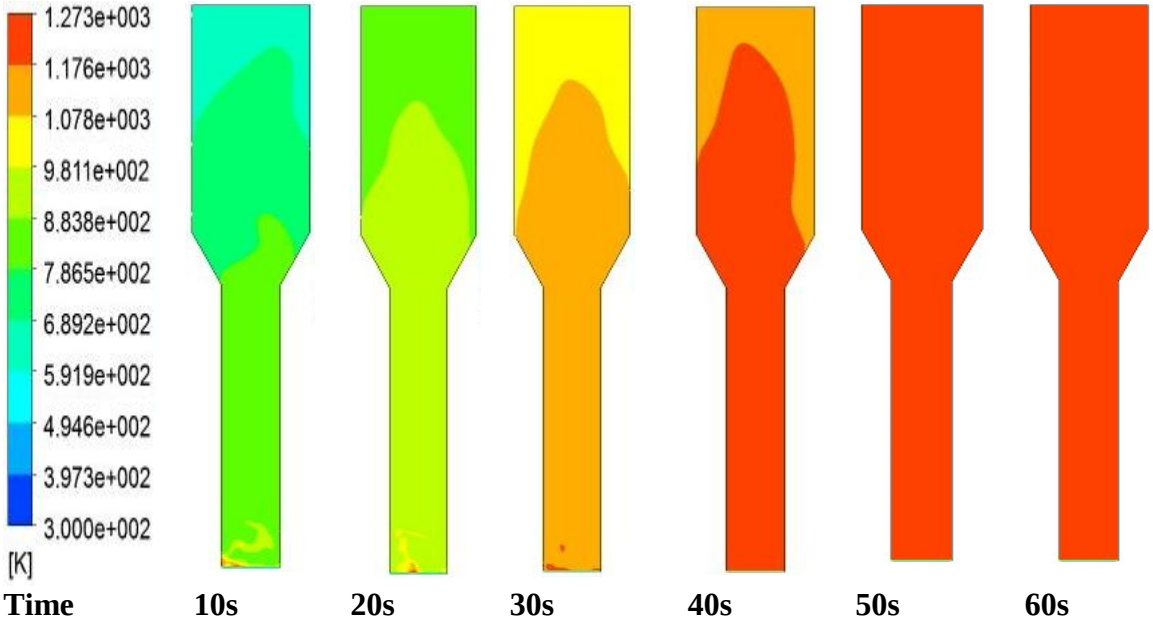
(vi) For rice husk

(G) Contour plots for temperature distribution for different biomass samples

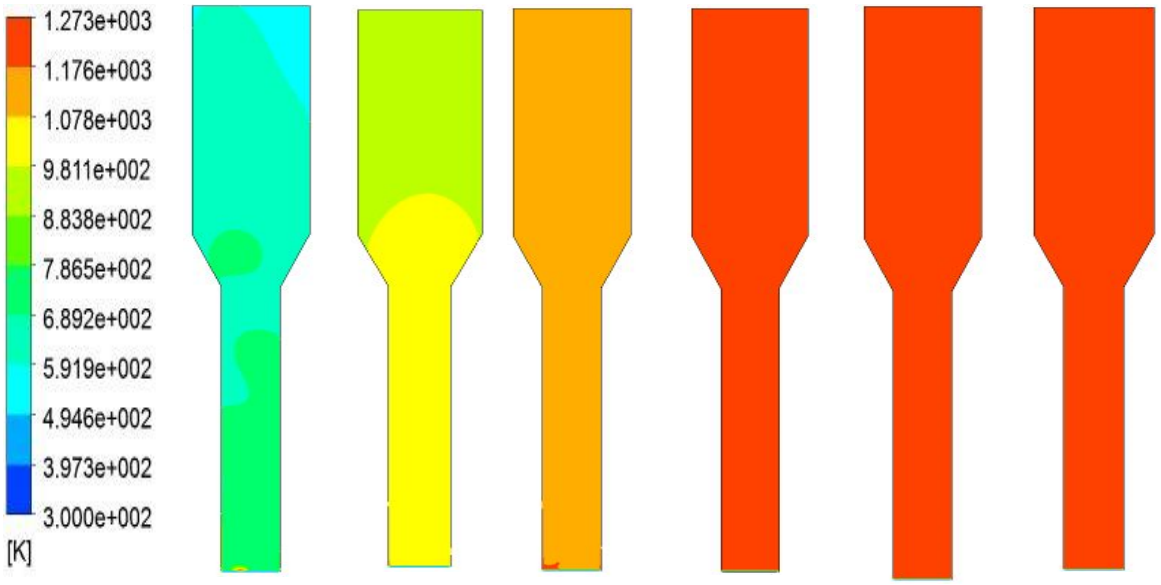
I. With 2D Simulations



(a) For coconut coir

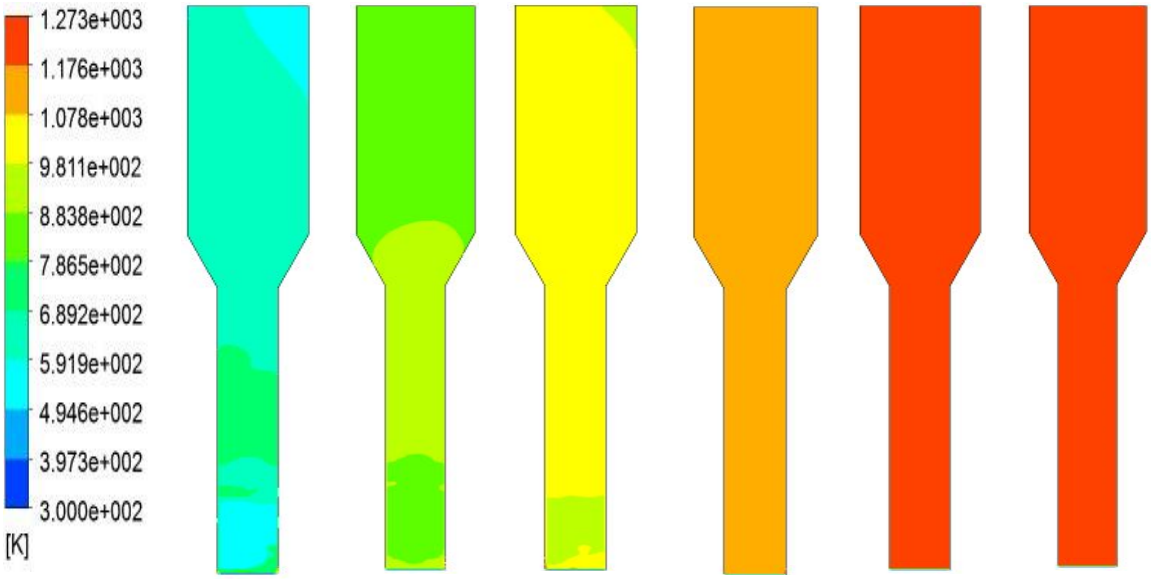


(b) For wood chips



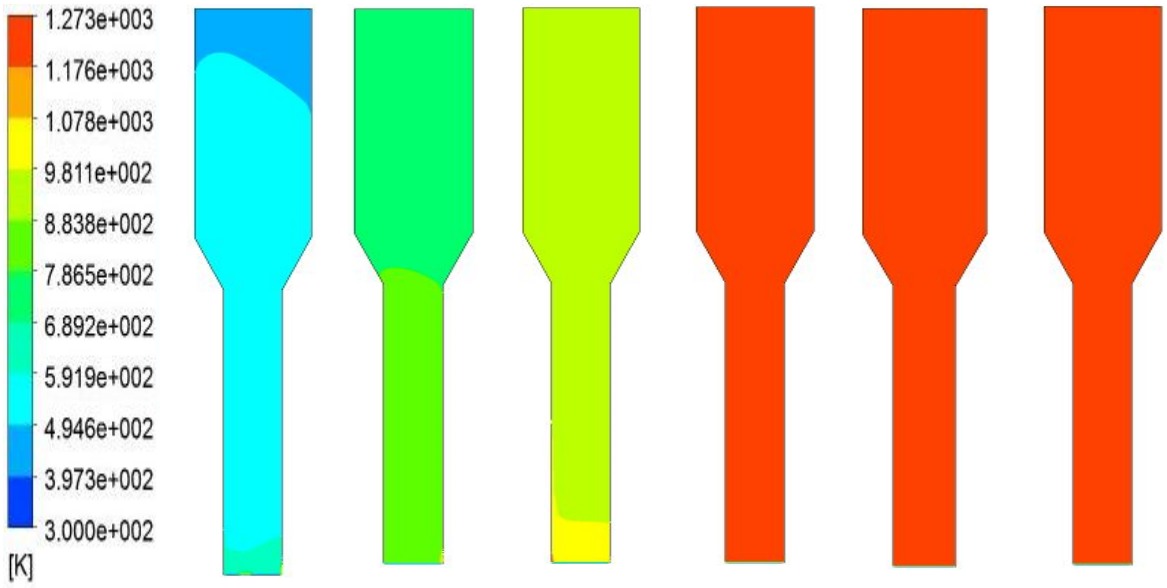
Time 10s 20s 30s 40s 50s 60s

(c) For rice straw

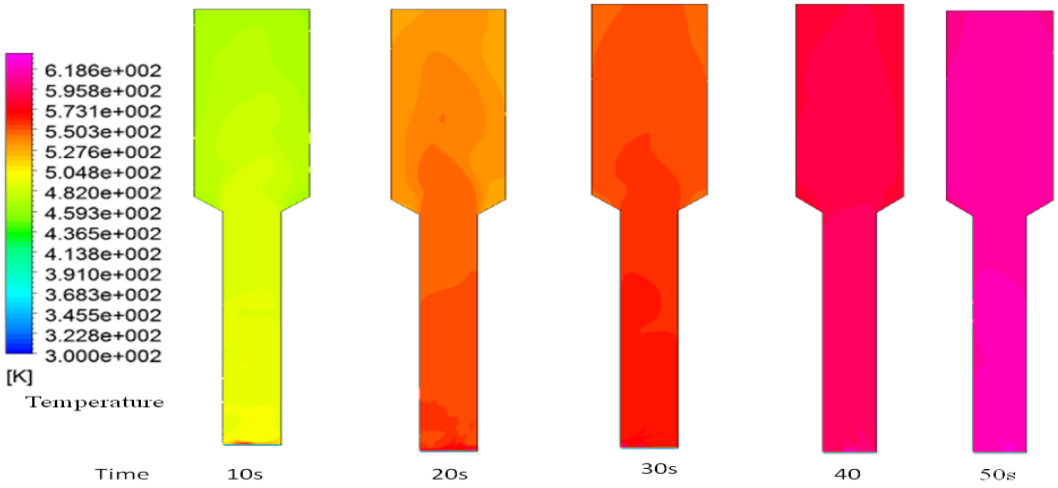


Time 10s 20s 30s 40s 50s 60s

(d) For sugarcane bagasse

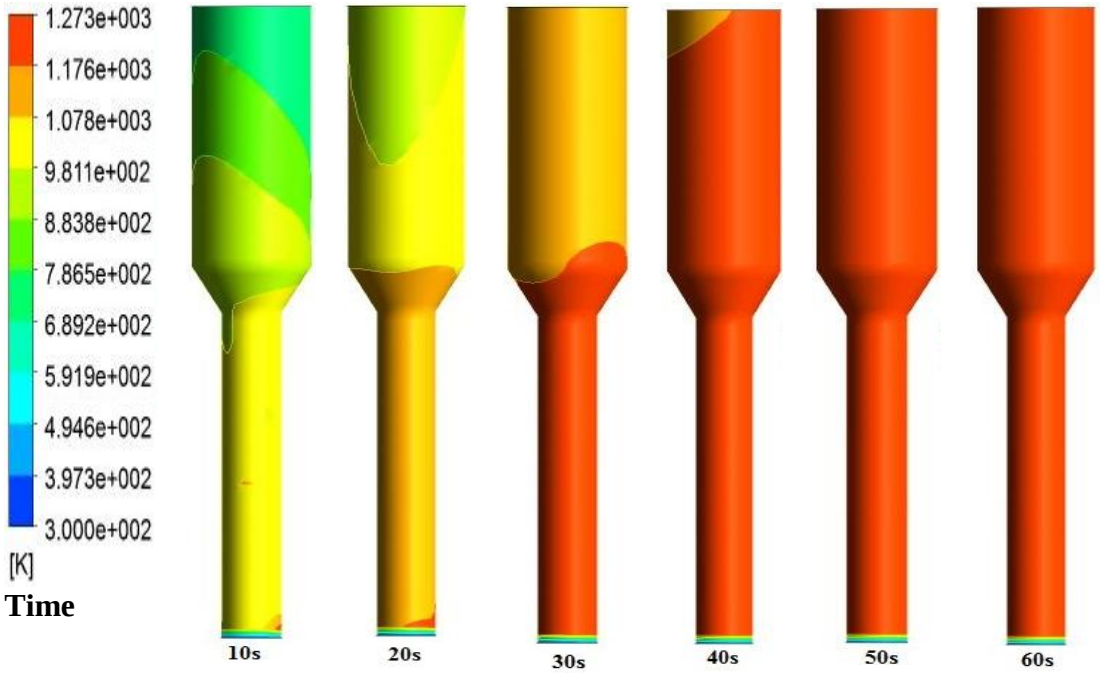


(e) For saw dust

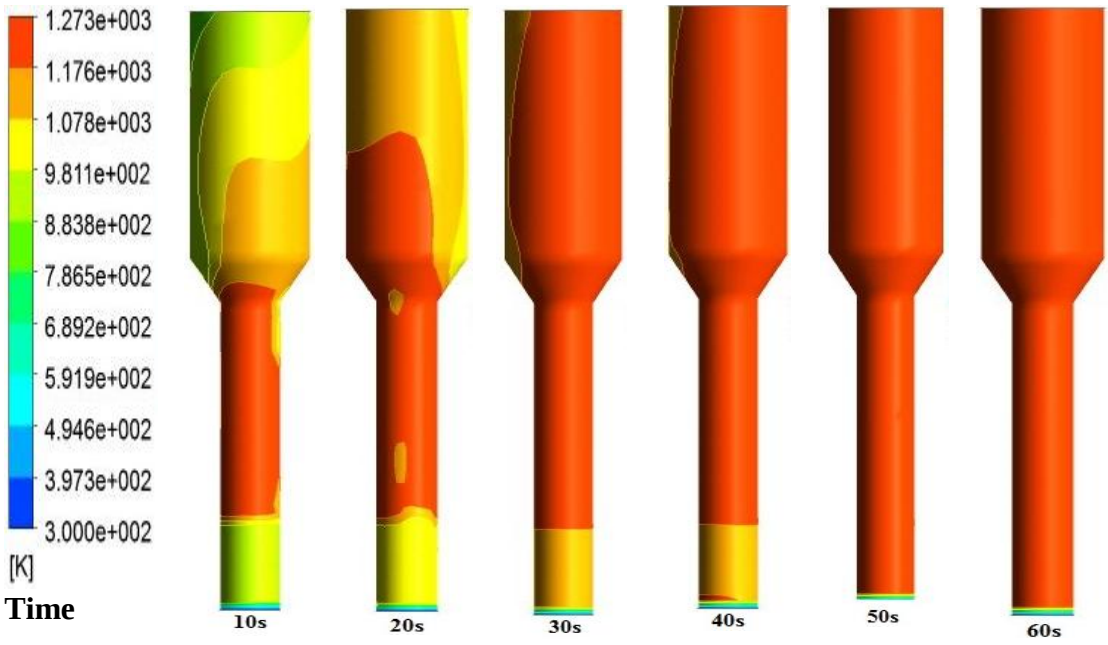


(f) For rice husk

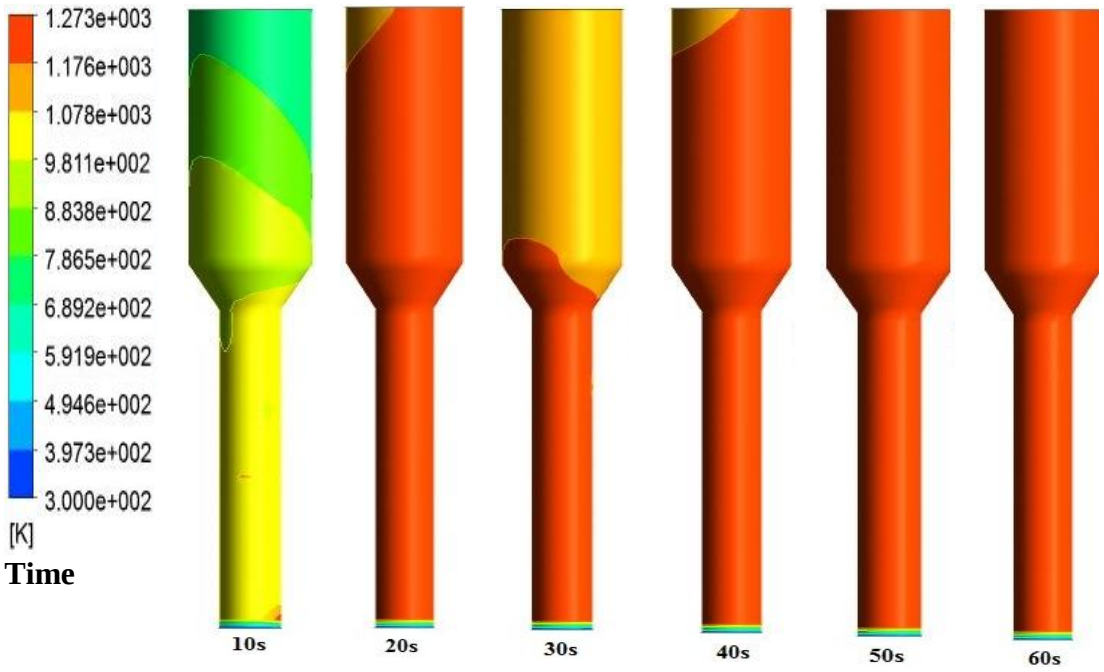
II. With 3D Simulations



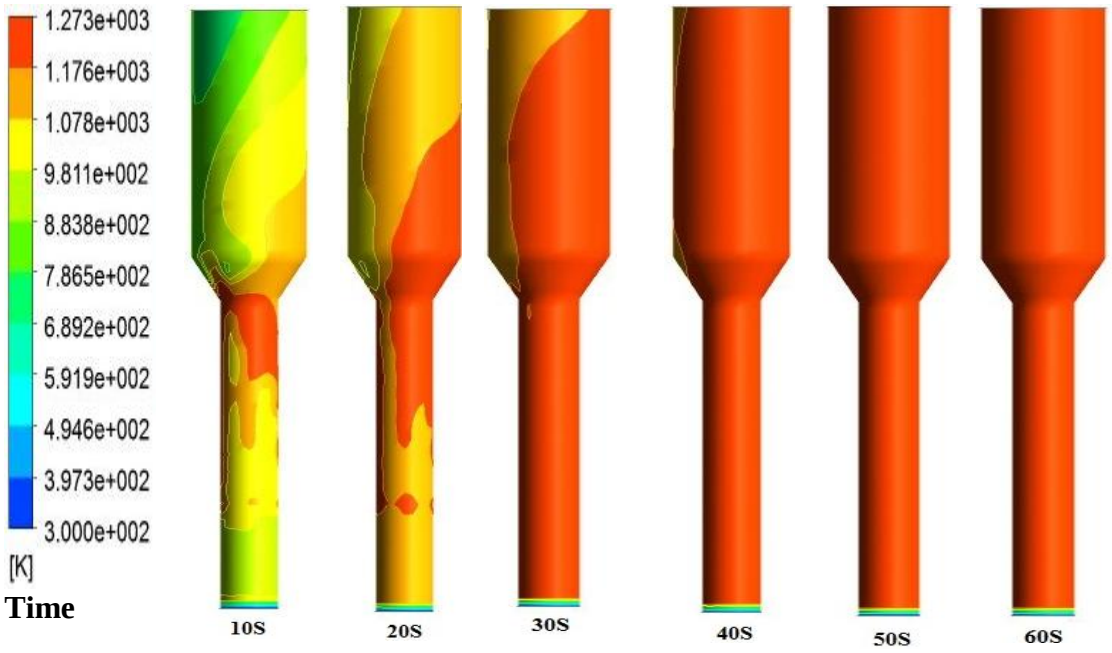
(a) For coconut coir



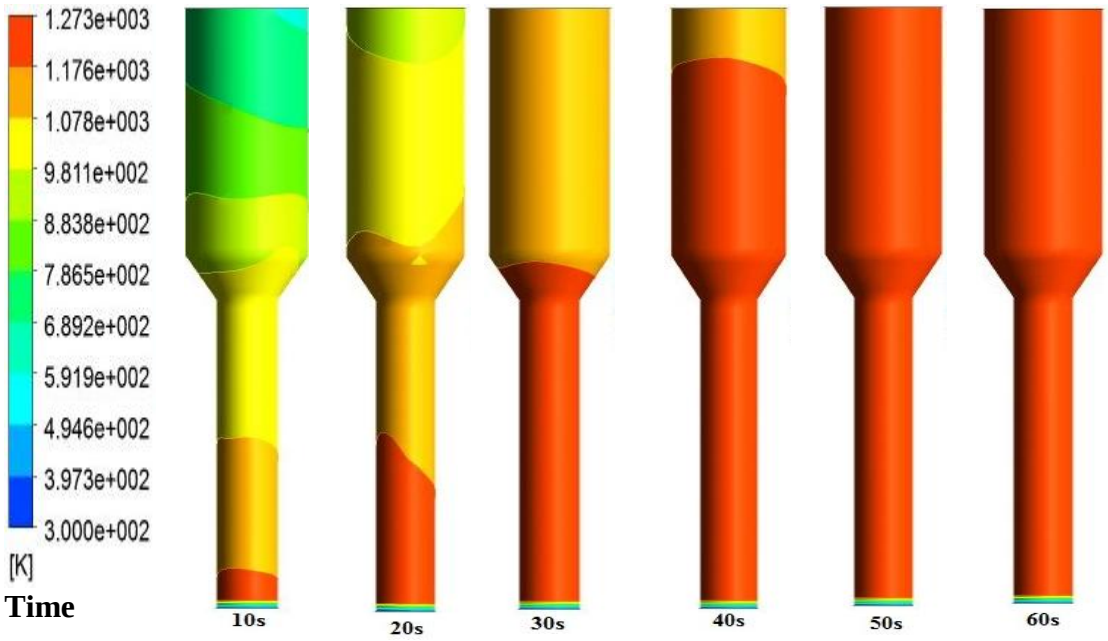
(b) For wood chips



(c) For rice straw



(d) For sugarcane bagasse



(e) For saw dust